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ADVANCED WASTE TREATMENT AND WATER REUSE SYMPOSIUM HELD AT PICK-CONGRESS HOTEL, FLORENTINE ROOM, CHICAGO, ILLINOIS ON FEBRUARY 23-24, 1971. VOLUME I. SESSION 1

Environmental Protection Agency
Cincinnati, Ohio

1974

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VOLUME I ADVANCED WASTE TREATMENT AND WATER REUSE SYMPOSIUM

PICK CONGRESS HOTEL
CHICAGO, ILLINOIS
FEBRUARY 23-24, 1971

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ADVANCED WASTE TREATMENT AND
WATER REUSE SYMPOSIUM

PICK-CONGRESS HOTEL

Florentine Room

Chicago, Illinois

February 23-24, 1971

Sponsored By

Environmental Protection Agency

and

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State Water Pollution Control Agencies of:

Illinois
Indiana
Michigan
Minnesota
Ohio
Wisconsin

ADVANCED WASTE TREATMENT AND WATER REUSE SYMPOSIUM

Session #1, Tuesday, February 23

Moderators: Staff Members, State Water Pollution Control Agencies - Illinois

9:00 A.M. Welcome

David P. Currie, Chairman
State of Illinois
Pollution Control Board

Introduction

Francis T. Mayo, Regional Director
Region V, WQO, EPA
Interim Regional Coordinator,
Region V, EPA, Chicago

9:20 A.M. Objectives of Symposium

Clifford Risley, Jr., Director
Office of Research and Development
EPA, WQO, Region V, Chicago

9:45 A.M. Concepts of Wastewater Reuse

F. M. Middleton, Director of Research
Advanced Waste Treatment Research
Laboratory, EPA, Cincinnati

10:30 A.M. Measurement and Forms of Nitrogen and Phosphorous

Edwin F. Barth, Chemist
Biological Treatment Branch, Advanced
Waste Treatment Research Laboratory,
EPA, Cincinnati

11:15 A.M. to 12:45 P.M. Lunch 2<

CONTENT

"Measurement and Forms of Nitrogen and Phosphorous"

Edwin F. Barth, Chemist, EPA

Additional Papers Included for Future Reference

"Why Advanced Waste Treatment and Water Reuse in the Megalopolis?"

Henry J. Graeser

"Gaining Public Acceptance of Waste Water Reuse"

Mrs. Jessie Rudnick

"Mineral Removal by Ion Exchange, Reverse Osmosis and Electrodialysis"

Franklin D. Dryden

"Renovated Waste Water Creates Recreational Lake"

Franklin D. Dryden and Gerald Stern

INTRODUCTION

I am honored to be a part of this cooperative venture co-sponsored by the Environmental Protection Agency and the States of Illincis, Indiana, Michigan, Minnesota, Ohio and Wisconsin.

We are witnessing a dramatic change - a virtual revolution in the shift of natural goals and priorities away from defense and space toward problems associated with our earthly environment. This shift will take on an accelerated rate very shortly, calling for not only a more precise definition of needs and priorities, but for a substantial reduction of the time gap between the development and technical demonstration of our agency's research, development and demonstration results, and their actual acceptance and implementation.

We have recently established a "Technology Transfer" program, which has, as its top priority goal, the translation of research results into modern, effective, concrete and steel structures staffed with adequately trained personnel. This symposium is a part of that program, a program designed to publicize the potential of presently available technology to reach all levels - the consulting engineers, city managers, decision makers, and the general public.

Francis T. Mayo

Francis T. Mayo
Interim Regional Coordinator, EPA
Regional Director, Region V, WQO, EPA

ADVANCED WASTE TREATMENT AND

WATER REUSE SYMPOSIUM

MEASUREMENT AND FORMS OF NITROGEN AND PHOSPHORUS

Edwin F. Barth
Advanced Waste Treatment Research Laboratory
U.S. Department of the Interior
Federal Water Quality Administration
Ohio Basin Region
Cincinnati, Ohio

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MEASUREMENT AND FORMS OF NITROGEN AND PHOSPHORUS

As our wastewater treatment goals increase in scope it becomes critical that good analytical control of processes be employed.

In order to implement and control nutrient removal processes we need a basic understanding of the chemistry of the elements nitrogen and phosphorus.

Phosphorus and nitrogen have many similar chemical reactions but there are some important dissimilarities.

A major difference between compounds of these two elements is that most of the phosphorus compounds we deal with in wastewater treatment are insoluble, while most of the nitrogen compounds are soluble.

In biological processes phosphorus is a conservative element. Loss to the atmosphere is not probable. In environmental situations phosphorus cycles between organic and inorganic forms. Microenvironments can have a great effect on availability of phosphorus.

The phosphorus procedure given in Standard Methods is for boiler water analysis and will not measure all forms of phosphorus present in wastewater. To measure treatment efficiency a total phosphorus procedure is necessary. Coupled with the total phosphorus assay it can be useful to filter a sample to determine the influence of liquid-solid separation processes.

Careful control of the final colorimetric assay conditions is important for accurate values and to avoid interferences.

With the increased interest in nitrogen in the environment the analysis for the various forms of nitrogen are assuming increased importance. The methods for organic nitrogen, ammonia nitrogen, nitrite, and nitrate are specific for each form. To completely account for nitrogen in a treatment system all these forms must be measured.

Preservation of samples is an important consideration in order to insure that loss or transformation to another species does not occur. Rapid freezing of samples for preservation is the most suitable method.

Automated methods are needed to collect sufficient data for evaluation. Careful attention is needed to insure that the electronic signal recorded is representative of the specific component desired.

The attached tables and figures will be used to present this topic.

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TABLE 1

WASTEWATER TREATMENT GOALS

1900-1920 HISTORIC

REMOVE SUSPENDED SOLIDS
REMOVE OXYGEN DEMANDING MATERIALS
TRANSFORM NH_4^+ TO NO_3^-

1920-1964 CONTEMPORY

REMOVE SUSPENDED SOLIDS
REMOVE BOD_5
PROTECT RECEIVING WATER FROM TOXICANTS
CONTROL COLIFORMS

1964-CURRENT TREND

REMOVE SUSPENDED SOLIDS
REMOVE OXYGEN DEMANDING MATERIALS
REMOVE PHOSPHORUS
TRANSFORM NH_4^+ TO NO_3^- OR REMOVE NITROGEN
PROTECT RECEIVING WATER FROM TOXICANTS
CONTROL COLIFORMS
POSITIVE CONTROL OF SLUDGES AND BRINES

Element	**Symbol**	**Valences**	**Usual form**
Nitrogen	N	3, 5, 1, 2, 4	Gas
Phosphorus	P	3, 5, 1, 4	Waxy solid
Arsenic	As	3, 5	Metallic solid

Compounds with similar properties

↑	↑	↑	↑
HNO_3	N_2O_3	NCl_3	NH_3
H_3PO_4	P_2O_3	PCl_3	PH_3
H_3AsO_4	As_2O_3	$AsCl_3$	AsH_3
Acid strength	Reactivity	Unstable nature	Stable nature

Similar compounds with different properties

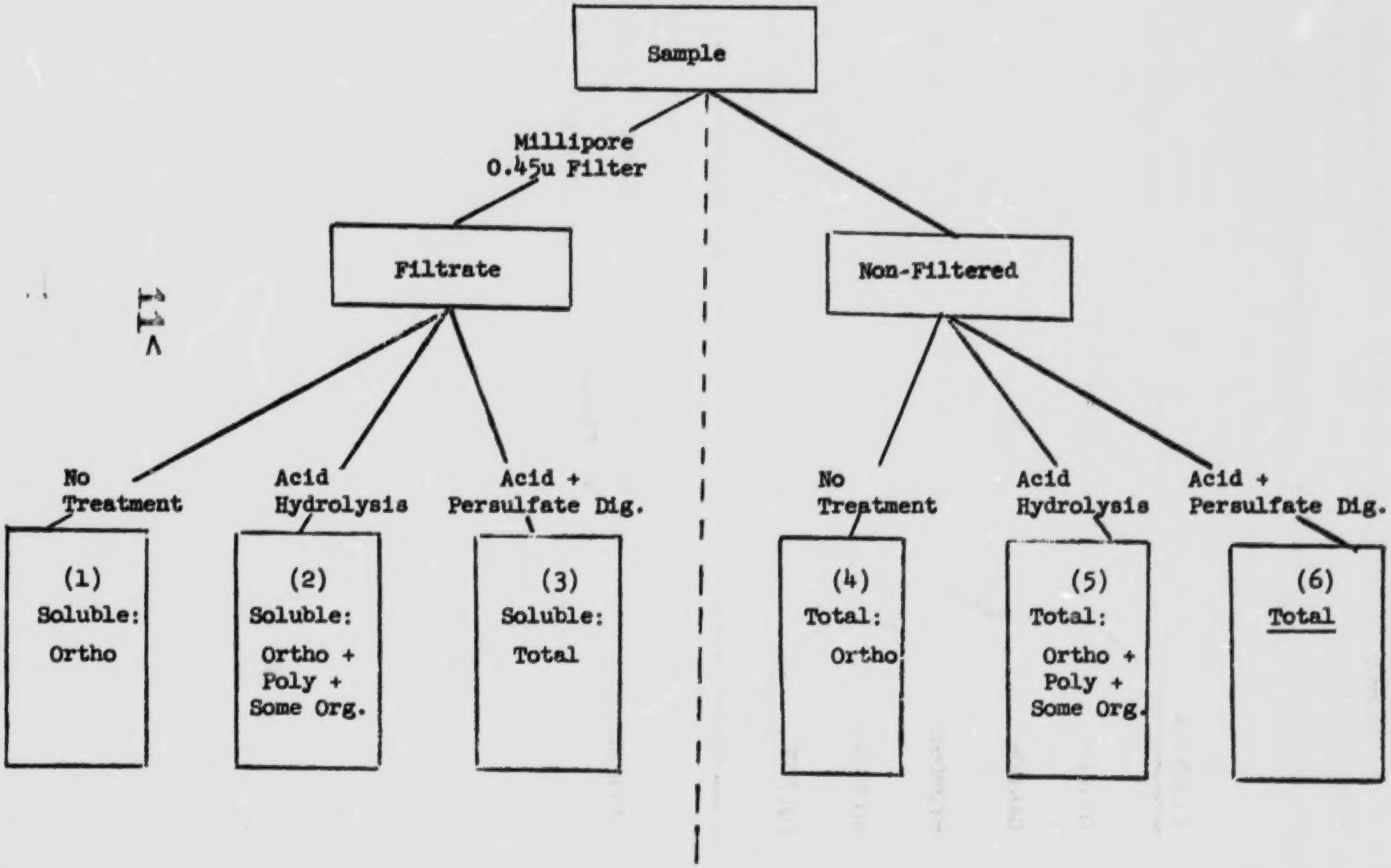
Soluble: $Ca(NO_3)_2$ $Al(NO_3)_3$ $Fe(NO_3)_3$

Insoluble: $Ca_3(PO_4)_2$ $AlPO_4$ $FePO_4$ $(NH_4)MgPO_4$

TABLE II. ELEMENTS AND BIOLOGICAL TREATMENT

ELEMENT	BIOLOGICAL REACTION	RESULT OF TREATMENT NET LOSS
OXYGEN	OXIDATION AND/OR REDUCTION	CO_2 CH_4
CARBON		
HYDROGEN		H_2 H_2O
NITROGEN		N_2
SULFUR		H_2S
		NO LOSS
PHOSPHORUS	NO OXIDATION NO REDUCTION	ORGANIC PHOSPHORUS ⇅ INORGANIC PHOSPHORUS

Determination of Various Phosphorus Forms


Sample
Millipore
0.45u Filter
Filtrate
Non-Filtered
No
Treatment
Acid
Hydrolysis
Acid +
Persulfate Dig.
(1)
Soluble:
Ortho
(2)
Soluble:
Ortho +
Poly +
Some Org.
(3)
Soluble:
Total
No
Treatment
Acid
Hydrolysis
Acid +
Persulfate Dig.
(4)
Total:
Ortho
(5)
Total:
Ortho +
Poly +
Some Org.
(6)
Total

Data Treatment	Phosphorus Components Determined
	A. Soluble Forms
1	1) ortho
2	2) ortho + poly + some org's.
3	3) total
2 - 1	4) poly + some org's.
3 - 2	5) most org's.
	B. Total Forms
4	6) ortho
5	7) ortho + poly + some org's.
6	8) total
5 - 4	9) poly + some org's.
6 - 5	10) most org's.
	C. Particulate Forms
4 - 1	11) ortho
(5-4) - (2-1)	12) poly + some org's.
(6-5) - (3-2)	13) most org's.
6 - 3	14) total

Forms and Measurements of Phosphorus

Phosphoric Anhydride	P_4O_{10} (usually calculated as P_2O_5)

Orthophosphates Found in Water

Phosphoric acid	H_3PO_4 below pH 2.2
Dihydrogen phosphate ion (monovalent)	$H_2PO_4^-$ from pH 2.2 to 7.2
Monohydrogen phosphate ion (divalent)	$HPO_4^{=}$ from pH 7.2 to 12.4
Phosphate ion (trivalent)	$PO_4^{\equiv}$ above pH 12.4
Dicalcium phosphate	$CaHPO_4$ from pH 6 to 8

(Tricalcium phosphate, $Ca_3(PO_4)_2$, does not form in water)

Hydroxyapatite	$Ca_5OH(PO_4)_3$ above pH 7-8
Fluorapatite	$Ca_5F(PO_4)_3$ above pH 7

Aluminum hydroxide adsorbs $HPO_4^{=}$ from pH 5 to 10

Aluminate ion $Al(OH)_4^-$ forms above pH 10

Ferric hydroxide adsorbs $HPO_4^{=}$ above pH 4

Magnesium hydroxide flocculates colloidal phosphates above pH 10-11

Polyphosphates

All polyphosphates hydrolyze slowly to orthophosphates

Sodium Trimetaphosphate	$Na_3(PO_3)_3$ -- a ring compound
Sodium Polymetaphosphate	$(NaPO_3)_n$ -- a long chain

(there is no justification for the name "hexametaphosphate")

Sodium Pyrophosphate	$Na_4P_2O_7$
Sodium Tri Polyphosphate	$Na_5P_3O_{10}$

Organic Phosphates

Esters and anhydrides	$ROPO_3^{=}$, $ROPO_3PO_3^{=}$
Phosphagens	$RNPO_3^{=}$

Reference

(1) Van Wazer, J. R. Phosphorus and Its Compounds. Interscience 1958.

PHOSPHORUS COMPOUNDS CLASSIFIED BY CHEMICAL AND SOLUBILITY RELATIONS(1)

Form	Water Soluble(2)	Insoluble(2)
1. Ortho phosphates $(PO_4)^{-3}$	Combined with monovalent cations such as H, Na, K, NH_4	Combined with multi valent cations such as $CA^{+2}Al^{+3},Fe^{+3}$
2. Poly phosphates $(P_2O_7),^{-4}(P_3O_{10})^{-5}(P_3O_9)^{-3}$ and others depending upon the degree of dehydration.	(a) as in 1 (above), but solubility decreases as molecular weight increases	(a) as in 1 (above) (b) multi P polyphosphates (high mol. wt.) including the "glassy" phosphates
3. Organic phosphorus R-P, R-P-R (3) (unusually varied nature)	(a) certain chemicals (b) degradation products (c) enzyme P (d) phosphorylated nutrients	(a) certain chemicals (b) cell mass, may be colloidal or agglomerated (c) soluble P sorbed by insoluble residues
4. Mineral phosphorus	(a) as in 1 (above), but solubility decreases for complex forms	(a) as in 1 (above) (b) geological P such as phosphosilicates (c) certain mineral complexes

(1) Table drawn from a lecture, "Determination of Phosphorus in the Aqueous Environment," by F. Ludzack, FWPCA Training Activities, Ohio Basin Region.

(2) Used to indicate usual state under common conditions.

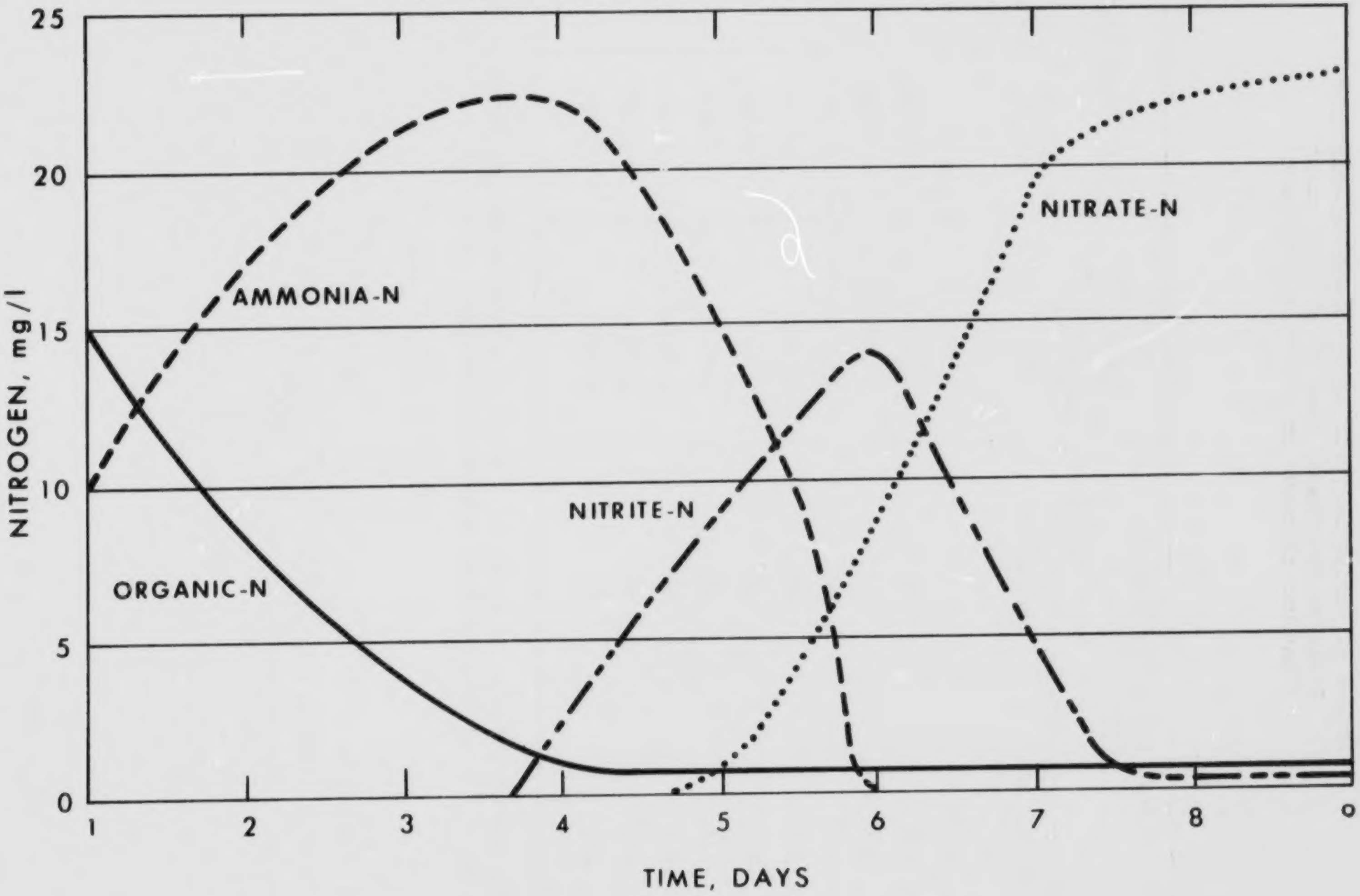
(3) R represents an organic radical, P represents P, PO_4, or its derivatives.

Effect of Varing Sulfuric Acid - Ammonium Molybdate Ratio on Heteropoly Blue Color Development
(Stannous chloride in glycerine reductant)

Acid Molybdate Reagent Composition		Normality* of acid in final 100 ml assay volume	Absorbancy (O.D.)	
Ammonium Molybdate gm/100 ml	Concentrated Sulfuric Acid ml/100 ml		Distilled Water	0.05 mg P Standard
2.0	5	0.072	> 2.0	> 2.0
2.0	10	0.144	0.805	1.00
2.0	15	0.216	0.260	0.580
2.0	20	0.288	0.036	0.475
2.0	25	0.360	0.004	0.415
2.0	30	0.432	0.000	0.365
2.0	35	0.504	0.000	0.345
2.0	40	0.576	0.000	0.252
2.0	50	0.720	0.000	0.060
2.0	60	0.864	0.000	0.012

*Calculated normality of H_2SO_4 in the 100 ml final volume produced by using 4.0 ml of the particular acid molybdate solution.

NITROGEN TRANSFORMATIONS



GPO 821-571A-2

Nitrogen Forms in 1:1 Mixture of High-Rate and Nitrified Effluents

Nitrogen Form	High-Rate Effluent	Nitrified Effluent	Denitrified Effluent	Oxidized Nitrogen Removal	Over-All Removal
Organic-N	5	2	3		
Ammonia-N	10	0	5		
NO_2-N	-	1 > 13	0.5 > 1.0		
NO_3-N	-	12	0.5	92%	
Sum	15	15	9		40%

VAN SLYKE REACTION

$$RNH_2 + NO_2^- \xrightarrow{acid} ROH + N_2\uparrow + H_2O$$

"WHY ADVANCED WASTE TREATMENT AND WATER REUSE IN THE MEGAPOLIS?"

--Henry J. Graeser

The "why" in the topic of my paper can really be answered by a few figures and in a few short words. Dallas has pursued a most aggressive program of water supply development since the early 1950's, having developed now four major reservoirs in the upper Trinity and Sabine watersheds. With the completion of these reservoirs a safe yield of nearly 360 million gallons per day has been developed, sufficient to carry Dallas County's expected population through the year 1995. Under planning is an additional reservoir on the upper Trinity which will extend available water resources to the year 2005. Planning studies are under way to look at available sources to the year 2020 and beyond. Evident in all of our projections is the fact that water costs are increasing both with distance and with inflation.

The Texas Water Plan, even with a reduced agricultural use, projects a water deficit of some 13 million acre feet per year by the year 2020. The deficit is in addition to a planned water reuse of some 8 million acre feet per year. This is quite pertinent and almost answers the topic question. The growing threat to the environment and depletion of pristine sources of water supply all testify to the imperative nature of the program in which you are engaged. Imperative as it is, however, other factors present themselves in this picture that must, of necessity, demand our primary attention. A recent article in Nations Cities listed the following data for growing water uses in terms of billion gallons per day:

In 1965 agriculture used	113 billion gallons per day
The 2020 projection is	166 bgd
Municipalities in 1965 used	24 billion gallons per day
The projection for 2020 is	74 bgd

Power generation in 1965 used	63 billion gallons per day
The 2020 projection is	414 bgd
Industry "self supplied" used	46 billion gallons per day
The 2020 projection is	211 bgd

Even with radical changes in America's life style with fewer and smaller automobiles, mass transit, better air conditioning, air pollution control and water resources planning which would involve reduction of return flows and water usage through thrift pricing and apartment dwelling, we face a formidable threat to both the environment and to the availability of water. Mr. George B. Maxey, Director for the Center of Water Resources Research in Reno, Nevada, said in the year 2000 the United States will have a population of 300,000,000 people. To provide jobs for these people and to maintain our standard of living will require 100% increase in mineral, fuel and power. He projected a 400% increase in water use by the year 2020. Economist, Robert Nathan, speaking at a national conference on public land policy and environment, using the same estimate of 300 million people in the United States by the year 2000 stated that "even if a law should be passed prohibiting women from having more than 2.11 children (the number to reach zero population growth), enough women are already born to assure a 275 million population by 2000."

The Texas Water Plan forecast a population of 30 million people for Texas in the year 2020. It is highly likely that Houston, Dallas, Fort Worth, and San Antonio can expect a lion's share of this population increase. It is fairly evident that appropriative water is a vanishing commodity. The future of any Texas city depends upon its developing and obtaining appropriative rights to meet this seemingly unavoidable future. These facts overshadow the interim problems of stream pollution. I use the term "interim" because the problems we now face in meeting stream standards based on recreational uses of our streams are obviously inadequate to speak to the question of a true water shortage In an area such as Texas where an evident water deficiency exists, part of everyone's water future will be reuse of return flow. The criteria that we have customarily used also pales into insignificance. Bacterial pollution becomes the easiest problem facing us and new criteria based on chemical content, heavy metals and organics will become the governing factors. Stream quality will be based on reusability rather than just recreational and ecological qualities. I can speak from experience that when you have no water, any water looks good. The question is - Can you drink it? - Can you use it?

Overshadowing our immediate problems of stream pollution is a growing competition for water. Agriculture, industry and municipal needs will vie for available supplies. Agriculture is facing this matter right now in that it cannot afford to pay the price for irrigation water which longer distances and pollution control place upon it. In any direction you look water reuse is in the picture. Furthermore, totally consumptive uses like agriculture have the cards stacked against them in double measure. Water used for municipalities and industry creates a hundred times the value in new product and in job opportunities than agriculture. In practical terms it simply means that industry and cities can out-bid agriculture for available water resources. Industry and domestic are non-consumative users. Too, as supplies grow and more and more of our population live in the urban areas, it is fairly evident that the needs of industry and the municipalities will be served. Big cities are helpless giants in the present situation. Their citizenry are, by and large, enthusiastic in supporting the environmental program. They have plenty of water, plenty of jobs, and plenty of money. But nearly any office holder will affirm that on an issue of wide personal interest, the urban population has the votes to swing the balance. On other issues they simply don't vote in proportion to the population. What I am saying is that if water becomes too expensive or, for that matter, pollution control becomes too expensive, if job security and standard of living are threatened because the environmental protection program has affected the ability of supplying fuel, fiber and power to a mushroomed population, the population can and will vote to reverse the trend even if it means loss of environment.

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Now we have the interest and attention of the population in both water resources development and environmental protection. This is what we have been promoting for years. The problem is over-reaction and single mindedness. Our task, then, is to set a wise course. We are at a dawn or first light stage where esthetic barriers and interim standards prevent a realistic look at our goals. "Interim standards" I say because present standards are being set without consideration of economic facts of life. "Esthetic barriers" I say because as long as pristine supplies are available no one wants to consider reuse as either a present fact in part or a future in whole. But I simply have no doubt in my mind that the future we face is one which gives environmental control a sound economic base through reuse and reclamation of our water resources. The future will require acceptance of reasonable compromises in the use of our resources to maintain our population standard of living along with preservation to the maximum benefit. Recreation, relaxation and meditation all must accept tolerable compromises. We cannot preserve all the wilderness just to enjoy it if it is at the expense of the well being of the large majority. At the same time, we must plan our cities, rural, and wilderness environments so as to provide the maximum of pleasure and beauty, yet continue it as a land of opportunity. Let us keep in mind that scarcity enhances rather than depreciates value. I remember vividly a small but beautifully designed park in New York City which nestled in the midst of crowded city streets and skyscrapers. The increased attention given to pedestrian malls, expenditure for landscaping and greenery in downtown city areas can bring an entirely different aspect to city living.

So, "why advanced waste treatment and water reuse?" - simply because we have about 20 years to gain the knowledge that we need to implement it safely. Our water pollution control program should be directed at the obviously ultimate goal rather than stopping short with a pleasing stream dedicated to recreation. In every instance the water is an essential economic commodity which is going to have to be used again and again if we are to maintain our nation's prosperity. Our research and our construction efforts have all stopped short of this ultimate goal, not from any lack of desire on the part of the federal and state agencies involved, but limited as always by funds and organizational responsibilities. The water we drink is quite different from the water we can play and fish in. While the dollars which have been spent in research seeking better methods for water pollution control will certainly have their impact in solving the ultimate problem of the water we drink, it has suffered by its inability to address the ultimate goal rather than the interim solution. The re-formation of the environmental agencies into a single agency is a step in the right direction in that now the Environmental Protection Agency can address itself to the whole problem as I am sure it has wished to do in the past. Likewise, the Bureau of Water Hygiene which has been restrained by a complete drouth in funds for research in drinking water will hopefully be funded and staffed in such a manner as to supplement the effort needed to meet the whole problem.

Consider the effect of a water utilization program where we could utilize return flow in the total water budget for an area. The regional plan for the North Central Texas Council of Governments envisioned five or six watershed systems. These six include two area sewage treatment plants below Dallas and a central sewage treatment plant on the East Fork below Lake Ray Hubbard and an extension of the TRA central interceptor system reaching all the way to Lewisville on the Elm Fork. If return flow were of such a quality that it could be considered part of the area water resources, discharges of drinking water quality could be returned to water reservoirs utilizing the dilution available and at the same time increasing the safe yield of these reservoir facilities. The savings in interceptor costs to divert waste water flows below impounding reservoirs would be saved, and these are some of the biggest expenses involved. Likewise, even though waste water treatment costs would increase several times, probably in the vicinity of 15¢ or 16¢, this is approximately what imported water from 100 miles away will cost. Thus, if a 100 million gallons of water were returned to the water sources of just the Dallas area, the savings in water supply development and importation would amount to $5,475,000.00 a year at 15¢ a thousand. This can be considered a savings only when you compare it to the present practice of returning flows to the streams for discharge to the gulf and admit stream standards will eventually dictate tertiary treatment on the Trinity River. The proposition whereby such water would be returned to reservoirs with dilution will also eliminate many of the stringent requirements in quality control which is implicit to direct reuse because a time lag is available for normal analytic tests for toxic concentrations which would be dangerous to the public health either for recreation or drinking water. The only problem, then, for this first step is purely esthetic.

It is interesting, too, that in such a scheme water which would have to be pumped to the reservoirs from waste water treatment plants below the impoundments could be scheduled only during periods of drouth when reservoirs fail to recharge from natural runoff. If we could only accept, then, the idea of reusing water, you are given an economic reason for very high degrees of more expensive treatment processes which would produce a drinking water quality into the streams. Other possibilities offer themselves, too, in that during flood periods or flood discharges from reservoirs a savings in treatment costs could be effected by eliminating some of the tertiary processes and at the same time eliminate the cost of pumping to storage. Under these conditions the normal 20 ppm B.O.D. would be more than adequate since the dosage of nutrients and pollutants both would be dispersed and intermittant. So, immediately before us is the economic incentive to tertiary treatment. The only barrier lies in improved technology to reduce the costs, improved process control and public acceptance.

In conclusion, let me simply mention some of the more obvious areas where we are deficient if we are to consider direct reuse. Primarily our technology for secondary and tertiary treatment of waste waters is still deficient despite the very fruitful research we have been prosecuting. There is a huge cost jump between conventional activated sludge or trickling filters to alternate methods of treatment. Despite all our research we still have not found more efficient processes at lower cost. Chemical and mechanical methods of purification create additional disposal problems. The biological process still reigns in attacking the problems of dissolved pollutants in waste water. Above all, we do not have the key to what makes the process work in all instances. This means we can't control the process or design it efficiently for all possible conditions. To put an economic base under water pollution control through water reclamation, we are faced with a degree of quality control never before considered necessary in waste water treatment plants. We are faced with a degree of control of what goes into the sewer never before considered necessary. Compounding this problem is a continued and increasing investment in plant facilities and collection system which we know with increasing uneasiness will not be adequate to the task before us and cannot be rectified readily because of the financial burden involved. We are committing our resources in many instances to inadequate solutions and these resources are indeed limited. We are going to have to come up with analytical controls that are "fail safe" if water is to be reused. We are going to have to stop overwhelming the problem with surplus capacity and begin to think in maximum utilization through process control coupled to reliable on-line instrumentation which will detect those parameters which affect treatment efficiency and the safety of the end product. As a minimum we are going to have to immediately initiate higher standards of house plumbing, house building sewers and collection system materials and workmanship. There must be radical changes in design and repair methods of collection systems to insure that these facilities exclude the entrance of extraneous flows which overwhelm and depreciate the efficiency of treatment processes. We must begin the almost financially impossible task of rebuilding the collection systems of our cities in such a manner as to exclude storm water infiltration. We must greatly expand and implement an industrial waste surveillance and control which will extend even to the household through exclusion of products which are dangerous to the ultimate goal of water reclamation and reuse.

But above all, we must keep our eye on the ultimate objective and that is reuse not just stream pollution control. We must and can grapple with the problems of the environment. We must address ourselves to the population problem and the increasing tendency of these new people to urbanize. We must solve the in-city environment problem if we are to live in the megapolis and that environment must be one which includes functional beauty in such a way as to inspire its citizens to perform efficiently and live constructively. We must think in terms of human benefits which are both economic and philosophical, because both are involved in our choice of priorities. We must not let pollution become a smoke screen to distract us from more immediate and essential challenges to mankind. It is a demanding part of the challenge - but it finds its solution through broader challenges essential to our well being and survival. The water we drink is one of those basic human needs.

League of Women Voters of the U.S.
1730 M Street, N.W.
Washington, D.C. 20036

January 12, 1971

GAINING PUBLIC ACCEPTANCE OF WASTE WATER REUSE

by
Mrs. Jessie Rudnick
Committee on Environmental Program and Projects
League of Women Voters of the United States

I want to say that I am delighted to have this opportunity to talk to you about waste water reuse because you are the professionals, the ones who can answer the questions the public is likely to ask about waste water reuse. You are the ones who will want to bring the answers to the attention of the public so that we can get on with the job of cleaning up our nation's water resources in order to make the best and fullest use of them.

The League of Women Voters has taken a strong position in favor of improvement in water quality. We have supported full authorized federal funding for sewage treatment plant construction, taking an active part in the Citizens Crusade for Clean Water which won an appropriation of $800 million for fiscal 1970 and $1 billion for fiscal 1971. We hope some of this money will be used to provide advanced waste water treatment, making use of techniques which have been developed in pilot plant and laboratory studies sponsored by the Federal Water Quality Administration.

To me this is an exciting prospect. I believe that the use of reclaimed water, even in the potable water supply, will be the "in" thing only a few years from now. Who doesn't want to be in the forefront of progress? Those of you who are planning now for waste water treatment facilities which will produce effluent of drinking water quality will be the heroes of your communities very soon - yes, even if it does cost more. I can hear it now: "My community is a better place to live than yours because we treat our sewage so thoroughly that we can even drink the effluent." "We don't pollute our river. It's safe to swim in." "Our children have a wonderful time at the city pool in the summer. It takes water right from the sewage disposal plant." "Our factories have cleaned up too. And, do you know, they've found that it doesn't cost any more to recover the minerals than to dump them into the river?"

There are two basic reasons why I make this prediction. One is that we are rapidly finding out, even in relatively well watered areas, that our supply of fresh unpolluted water is not inexhaustible. Another reason is that we can no longer contemplate disposing of our wastes indefinitely by just dumping them into the nearest body of water. The concept of the closed system, with all wastes reclaimed as much as possible and only makeup water added, is one we can and must achieve if we are to save our environment.

In preparing for this occasion, I asked several of my friends how they would react if they were told that from now on reclaimed sewage effluent would be returned to the city water system. The reply was either a stunned silence or a grimaced "ugh." However, in every case, after this first startled reaction, a further discussion revealed that the person I talked to would be quite willing to accept such an action, provided he could be convinced that the effluent had been treated in such a way that it was safe to drink. Therefore, I want to talk to you today about some of the questions I believe people will want you to answer when you present to your community proposals for waste water reuse.

I feel sure the first question will be: "Is it really possible to clean up our sewage so that the effluent will be safe to drink?" The answer, as you know, is: "Yes, the technology has been developed to do just that." However, I say this with some reservations because even though we have pilot plants which can produce potable water from domestic waste water, there still is, to my knowledge, no full scale system in operation in this country which actually returns potable water directly to the water supply system. Other countries are reusing waste water successfully. For example, the city of Windhoek, in Southwest Africa, a water short area, has a sewage treatment system which provides about one third of the total potable water for the city water supply.

The Summary Report - Advanced Waste Treatment - July, 1964-July, 1967 issued in 1968 by the Federal Water Quality Administration, then the Federal Water Pollution Control Administration, makes surprisingly interesting lay reading, at least to those of us who are concerned about the deterioration of our water quality. We've learned that every stage of advanced treatment of domestic sewage has been tested experimentally in this country and that there are several different ways of achieving an effluent of drinking water quality which may be tailored to the quality of the raw sewage. There are pilot plants in Lebanon, Ohio; in Pomona, California; and in Washington, D.C., where new methods of treatment are being developed. In addition, there are many other places where advanced methods of waste water treatment are being tested. In every case, though, the system stops short of returning the effluent directly to the potable water supply.

The Whittier Narrows plant in California recharges an aquifer. The Pomona, California plant effluent is used for irrigation. Effluent from the South Lake Tahoe water reclamation plant is piped over a ridge to a new man-made recreational lake. The Lake Tahoe effluent meets USPHS drinking water standards, yet it is not returned to the lake. Why? What keeps us from taking that logical next step of reusing the water directly? There appears to be a widespread belief that the public will not accept such a step. I believe that the public will accept such a step when given the right information. I hope that soon the effluent from many pilot and experimental installations in this country will be returned to the potable water supply. And when this does happen, I hope the event will be widely publicized so that many people will hear of it. I hope any of you who are there will be glad to publicly drink a glass of water from one of the advanced treatment plants so that everyone will see that it is drinkable. Actually, you may find that the water tastes flat because it lacks some of the contaminants you are used to.

In order to combat unfavorable rumors and untruths, the public needs basic information. Admittedly it may not be easy to explain some of the more sophisticated water treatment processes to the general public but I believe the public would certainly like to know what can be accomplished. For example, one can hardly pick up a newspaper or magazine today without seeing a reference to the damaging effects of excessive amounts of phosphates in waste water but few people seem to know that the phosphates can be removed from the waste water. As with the foam from non-biodegradable detergents, the solution to excessive phosphates may lie in changing the composition of detergents but I'm sure the public would like to know what alternatives are available.

The League of Women Voters believes that the cost of cleaning up industrial waste water should be considered a cost of doing business, although limited federal assistance may be necessary in some cases at the beginning. We should not like to see the possibility of waste water reuse jeopardized because of industrial contaminants. We hope that the job of cleaning up our water supplies will not be slowed down because domestic and industrial wastes have been combined in the past.

Once the public is convinced that waste water reuse is technologically feasible, or possibly even before, the second question will be raised. "Do we really need to go to the trouble and expense of recycling our waste water?" Again I believe the answer is "Yes," and again the reasons are the same. We are running out of new supplies and we shouldn't be allowed to dump our wastes into what may be someone else's water supply. I think a fairly large segment of the public is already concerned about the deteriorating quality of our water supplies in many places and I am also sure it is apparent to some people that we are already recycling our water. As communities grow in size and number, it becomes increasingly unlikely that the waste water from the town upstream will have purified itself by the time it reaches the fresh water intake of the town downstream. If the water comes from wells, we are finding that often the supply will not meet the increasing demands. New wells may be drilled but the underground supply is not inexhaustible.

The question is, "At what point do we decide that the water in our fresh water source is no better than the water from our waste treatment plant?" Hopefully we will be giving advanced treatment to our sewage and recycling it before that point is reached. Some of the advanced waste treatment techniques are identical with those used to purify intake water; lime coagulation and sedimentation, for example. There should be some possibilities here for an integrated system.

In a wry sort of way our prodigality may be our salvation. We've been so prodigal about throwing away our used water that we have quite a lot to salvage. Many coastal cities have been dumping their waste water into the ocean, often with little treatment. Even in well watered areas this amount of clean water would make a significant addition to the city water supplies.

It is becoming increasingly hard to find adequate sources of new fresh water in the west especially. Expensive water transport systems may have to be built to provide the water needed if we do not turn to recycling. Alternatively we may need to curtail agricultural use of water in order to provide industrial and domestic supplies if we do not recycle the water.

When we are faced with decisions like these, we come to the third question which will be asked by the public. "Is it economically feasible to recycle our waste water?" The cost of tertiary treatment facilities will be high, possibly as much as three times that of providing secondary treatment. Since we have fallen behind in so many places in providing even adequate secondary treatment, the cost of providing treatment which will enable cities to recycle their water will indeed be high. However, we are already paying a high price for pollution in our rivers, oceans and lakes. This is a price we pay in disease, in loss of recreation, in the loss of both commercial and recreational fishing opportunities, and in loss of property values. We must balance these costs against the costs of cleaning up the water.

Balancing these direct and indirect costs against the cost of tertiary treatment may easily result in a balance in favor of reuse. Even the cost of one new well may be more than the cost of recycling an equivalent amount of water.

The city of Santee, California found that it was cheaper to give tertiary treatment to its sewage than to build a new sewage treatment plant. In addition, it reaped the indirect benefits in water-oriented recreation. Downstream users find the treated water preferable to water which comes from the Colorado River in open aqueducts because it has fewer dissolved salts.

The use of bottled water is increasing in this country, primarily because of deteriorating quality of water supplies. This is certainly an expense people would prefer not to have.

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This, then, is the information I think people want before they can be convinced that waste water should be recycled. They want to know that it is possible to clean up waste water so that it can be recycled. They want to know that it is necessary to do so, and they want to know that it may be cheaper to do so, when all expenses have been considered.

There still remains the question of how to get started. An astonishing number of people still don't really care where their water comes from or where it goes after use, as long as it is free from objectionable odor, taste or appearance when it arrives and as long as it disappears after use. How do we persuade the public to ask the questions I have discussed? I think the initiative should come from you. This is because you know the needs and the possible solutions, but I hope you will enlist the aid and support of concerned citizens groups such as the League of Women Voters, local conservation coordinating organizations and the Izaak Walton League.

Let us suppose you have surveyed the needs of your own community and have prepared a plan for water reuse. You send a copy to the president of the local League of Women Voters. The League water resources chairman reads it, decides that it is within the scope of its program and prepares to investigate the plan in order to present it to the membership for their opinion and perhaps approval. The League member will need background so she will ask questions. She will want to know where the city water supply comes from, why it will not be adequate for future expansion, what its quality is at present, how it is treated before use. She will want to know what happens to the water after it is used, how adequate are the sewage disposal plants, both in number and in quality of treatment. She will want to know costs, the cost of present supply and disposal and the projected costs. She will want to know whether our environment is being adequately protected now and whether the proposal will also protect the environment.

I hope you will tell it as it is, even if your town has fallen behind and is not meeting present standards of good treatment. Once League members have the facts and have reached consensus, they will help you alert the public and put pressure on the proper public officials. When all this has been done, I am sure the concept of waste water recycling will be readily accepted.

Together we should be able to achieve the support for waste water reuse which will be so important in the years to come. I hope that within my lifetime I shall see a real change from the present increasing pollution of our water supplies to rapidly decreasing pollution, with all the benefits to life on this planet that such a change implies.

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MINERAL REMOVAL BY ION EXCHANGE, REVERSE OSMOSIS AND ELECTRODIALYSIS

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Introduction

The purpose of this paper is not to argue whether or not demineralization of wastewater will some day be practiced, but to review some relevant research findings and cost estimates which may influence when and where such demineralization will occur. There are, of course, large quantities of wastewater in California with dissolved solids concentrations well above what might generally be conceded as acceptable limits. Such waters will probably not be reused unless they have been demineralized. However, there are many more situations in which wastewaters have marginal dissolved solids concentrations of slightly above or below 1,000 mg/l. In discussing waters of this quality with water supply agencies there appears to be considerable interest in reducing the dissolved solids to more desirable levels. This situation, which can be described as removing the increment of dissolved solids added to the initial water supply by one-time use in the community, seems to be the most probable application for demineralization in the near future, and most of the research work and cost estimates contained in this paper have been developed on that assumption.

The Sanitation Districts have been involved in pilot scale evaluations of demineralization processes since 1965. The work has been conducted jointly with the Federal Water Quality Administration at the Sanitation Districts' Pomona Water Renovation Plant. A portion of the 10 mgd activated sludge plant effluent is diverted to the Research and Development Facility shown in Figure 1. In addition to the various pilot plants, the research facility includes office and laboratory space for five engineers and seven chemists working on various studies. At the heart of this research facility is a 300,000 gallon per day, four-stage, downflow, packed-bed, granular activated carbon adsorption column. For the past five years this system has received typical secondary effluent from the Pomona plant and reliably produced a high quality carbon filtered water which has also served as the feed water for the demineralization processes. Average quality data for the activated sludge plant effluent and carbon adsorption product water are shown in Table 1.

Since all of the demineralization processes investigated at Pomona involve membranes or resins which are subject to fouling in the presence of high concentrations of organic matter, activated carbon adsorption is at present considered an integral part of these demineralization schemes.

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The cost of activated carbon treatment is estimated at 9¢ per thousand gallons for a 10 mgd facility, which must be added to the demineralization cost to arrive at the total cost estimates for each of the three processes presented at the end of this paper. The three processes -- reverse osmosis, electrodialysis, and ion exchange -- are presented in decreasing order of total estimated cost based on the studies to date.

Reverse Osmosis

Although the reverse osmosis process is the newest technological development of the three processes studied at Pomona, it was the first to be piloted there. There are many ways in which the principals of reverse osmosis can be converted to hardware and there are at least five or six companies with different versions of the process. Three of these have been piloted at Pomona but only two have been operated for sufficient length of time to permit a reasonable evaluation of their performance and economic potential. In the foreground of Figure 2 there is a 12,000 gallon per day unit manufactured by the Gulf General Atomic Corporation, and on the left edge of the slab a 3,600 gallon per day unit manufactured by Universal Water Corporation.

The Gulf General Atomic (GGA) unit features a spiral wound membrane packed into high pressure, 4" diameter pipes arranged in series. The configuration of the membrane module is illustrated in Figure 3. Each module is comprised of a wound up sandwich comprised of two layers of membrane for the "bread" around a "filling" of membrane support material. A brine side spacer separates layers of membrane as the membrane is wound around a low pressure product water tube at the central core. The brine stream passes longitudinally down the membrane through the spacer under an initial pressure of 465 psi. Water molecules pass through the membrane and travel spirally through the backing material to the low pressure product water collection tube at the center. Each module is 3' long and contains 50 square feet of membrane surface area. The unit shown in Figure 3 has a total surface area of 1,350 square feet and a flux rate of slightly under 10 gallons per day per square foot.

The Universal Water Corporation (UWC) unit features a more typical reverse osmosis design. The basic component of this system is a half-inch diameter perforated plastic tube which serves as the support for the membrane which is contained within the tube. Eighteen of these tubes are connected in series and housed within a 4" diameter product water collection pipe. The feed water flows through each tube at an initial pressure of 600 psi. The 4" diameter modules are connected in a combination of series and parallel arrangements to achieve the desired flow rate past each membrane. This unit contains 225 square feet of membrane area and has a flux rate of approximately 15 gallons per day per square foot, which is substantially greater than that of the GGA unit, even when accounting for the difference in operating pressure.

Although the present state of development of the reverse osmosis process results in the highest cost estimate of the three processes considered at Pomona, it is also the process with the greatest potential for technological improvement. Significant improvements in either flux rate or membrane life could appreciably lower cost estimates. In order to provide reasonably reliable estimates for the present state of technology, the Sanitation Districts are operating both the GGA and UWC units on extended runs to determine the life of the membranes and the rate of change in flux rate with time. The performance of the two units is compared in Figure 4. At this time the GGA unit has been onstream about 9,000 hours, or approximately one year, compared to about 5,000 hours, or seven months, for the UWC unit. These runs are believed to be the longest continuous runs of a reverse osmosis system for the demineralization of reclaimed water. Even so, it will require another year of operation of the GGA unit to confirm the two-year life expectancy which has been assumed for purposes of the cost estimate on this process.

The difference in flux rate between the two units is substantial and apparently represents a difference in the state of technological development of membrane production by the two companies. The rate of decline in flux rate has been identical for each unit and has been empirically found to follow a straight line when plotted against time on a log-log scale, as shown in Figure 4. During the course of the life run the influent water has been automatically maintained at a pH of 5 with sulfuric acid. There is a daily flush with a mixture of air and tap water and a weekly flushing with an enzyme detergent. Whether these routine operating procedures must be increased or can be reduced without affecting membrane life still needs to be determined.

Data on performance of the GGA unit is presented in Table 2. Average water recovery during this period was 75%, although it is anticipated that for a large installation a more advantageous flow pattern would permit recoveries of up to 85 percent. Because of the general interest in nutrient removal, Table 2 includes data on ammonia nitrogen, nitrate nitrogen and phosphate, as well as on TDS. It should be noted, however, that when these processes are used to remove the increment of TDS added during one municipal use of water, nutrient levels will not be reduced sufficiently to inhibit eutrophication problems. This process has maintained over 90% TDS reduction for an extended period and has the advantage of similarly high removal of COD materials. It is even more selective for the removal of phosphate but has only achieved about a 50% removal of nitrate nitrogen.

Electrodialysis

The 15,000 gallon per day electrodialysis unit constructed by Ionic Inc. is shown in Figure 5. A rectifier in the gray chamber in the foreground provides a 70 volt DC output across the membrane stack, just visible on the left. The membrane stack is 18" x 20" x 12" deep and contains 65 anion and cation permeable membrane pairs and their respective spacers. The flow diagram is illustrated in Figure 6. A flow of approximately 12 gallons per minute in the feed stream is split with 10 or 11 gallons per minute passing through the dilute stream and becoming product water as the cations and anions migrate into the concentrate stream. The remaining one to two gallons per minute of the feed stream is mixed with 10 to 11 gallons per minute of brine recirculation flow to provide equal flow rates and consequently equal pressures on both sides of the membranes. A pH of 3.5 in the concentrate stream is maintained automatically by the injection of sulfuric acid.

The percent conductivity removal during the first three months of operation of this unit is presented in Figure 7. Decreases in percent conductivity removal are closely related to decreases in amperage across the unit. The unit is designed to operate at 4 amps and generally achieves a dissolved solids reduction of about 34 percent. On the average, conductivity removal has been relatively stable. Most of the fluctuations noted on the graph are related to the changing character of the TDS of the water rather than to changes in the membrane itself. Routine operating procedures to maintain high performance include flushing with a sequence of an enzyme detergent, tap water, and acid every second day with the voltage shut off. This has permitted at least four months of operation without requiring manual cleaning of the membranes. The relative cost of more frequent manual cleaning vs. routine flushing has not yet been determined. Performance data from a previous period of operation is shown in Table 3. The process is more selective for nitrogen compounds and less selective for phosphate than for average TDS removal. This process has the disadvantage of being relatively ineffective on soluble organic material, as represented by the 15% removal of COD.

Ion Exchange

The ion exchange pilot plant shown in Figure 8 was originally supplied by Infilco Corporation with its own pretreatment unit. However, the pretreatment unit was unreliable and the cost of pretreatment was greater than the savings in ion exchange costs, which resulted in abandonment of the pretreatment portion and operation of the ion exchangers with carbon filtered water. A simplified flow diagram is presented in Figure 9 for this strong acid-weak base ion exchange process. The unit operates at 2.5 gallons per minute for a 16-hour period each day and produces about 2,400 gallons of product water per day. The primary cation exchanger contains 8 cubic feet of resin and the secondary cation exchanger 3.2 feet of resin, with each of the anion exchangers containing 1.6 cubic feet of resin. Once a day the last three exchangers are shut off and the primary cation exchanger is backwashed with feed water. Once a week each of the other exchangers is shut off in sequence and backwashed with its normal feed water.

A simplified diagram of the regeneration process is presented in Figure 10. Sulfuric acid is used to regenerate the cation resins and ammonium hydroxide is used to regenerate the anion resins. The regenerates are discharged to waste and the resins are given a product water rinse.

Comparison of the feed and product water TDS for 80 days of operation is presented in Figure 11. The influent TDS averaged slightly over 600 mg/l, and the product water TDS averaged about 70 mg/l. Ion exchange systems are generally used in industry to produce extremely low concentrations of dissolved solids. For demineralization of reclaimed water, however, substantially higher TDS values are not only permissible but desired to reduce costs. By reducing the degree of regeneration of the resins and operating beyond the normal breakthrough point for industrial water quality, it has been possible to reduce the cost of regenerating the resins while producing a very acceptable demineralized product water.

By manipulating the end points, as previously described, the ion exchange system has accomplished an average TDS reduction of 87% at an average water recovery of 87%, as shown in Table 4. COD removal was much better than expected in this process and phosphate removal is outstanding, although this ion exchange system is less selective for ammonia nitrogen.

Cost Comparison

Tables 5, 6 and 7 present comparative cost estimates for the reverse osmosis, electrodialysis and ion exchange processes based on producing 10 million gallons per day of product water. The reverse osmosis estimate in Table 5 still involves many assumptions. First, the capital cost has been provided by the Gulf General Atomic Corporation and is believed to be conservative, but until plants of this type are constructed on a large scale, reliable production costs will not be available. The cost of chemicals is surprisingly high in this estimate and is largely represented by the cost of pH adjustment and enzyme detergent flushing. There may be some possibility of reducing this cost in the future. The item for "maintenance materials" is figured at 5% of the depreciable portion of the capital investment on all estimates and is a relatively minor portion of the total cost. Power is naturally a substantial figure for a reverse osmosis system and cannot be reduced. The largest item is membrane replacement which is one of the reasons membrane life and flux rate are the points where major emphasis on improving technology must be placed. The effect of changing either of these variables is illustrated in Figure 12. A two-year life and 10 gallons per day per square foot flux rate has been used in the previous cost estimate. Accomplishment of a three-year membrane life and an increase of flux rate to between 15 and 20 gallons per day per square foot would make a 1/3 reduction in the total cost of this process.

At first glance the estimate for electrodialysis in Table 6 looks relatively low. However, this cost, as opposed to that of reverse osmosis and ion exchange, does not have the potential of being reduced by blending since the process only achieves about a 1/3 reduction in TDS. Furthermore, it is felt that the capital cost of the process as well as the operating cost are relatively well known, with only minor possibilities of making slight improvements in either chemical usage or labor costs in the future.

The cost of ion exchange, as presented in Table 7, appears to be the most reliable estimate of the three processes, although the cost of regenerants is related to the particular combination of ions included in the overall dissolved solids of the feed water. Furthermore, this process

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is the most sensitive to waters of variable or high dissolved solids content because the cost of regenerants is directly related to the number of ions removed. With regeneration costs amounting to 16.2¢ per thousand gallons out of the total estimated cost of 22¢ per thousand gallons, it is apparent why most of the effort has centered on improving the efficiency of utilization of regenerants. Resin replacement is not a particularly large cost item, although the 20% per year figure is considered somewhat conservative.

In Figure 13 the total cost of the three processes are compared under equivalent conditions for an influent TDS range from 500 to 1,500 mg/l. In each case it has been assumed the influent TDS is reduced by 33% through demineralization. For the electrodialysis unit 33% is approximately the reduction obtained in a single pass. However, both the reverse osmosis and ion exchange units provide 90% TDS removal, hence the cost of these processes have been adjusted by assuming a mixture of demineralized and non-demineralized carbon effluent to produce 20 mgd of the desired TDS product concentration. Though these stimates include the 9¢ per thousand gallons cost of carbon filtration, they do not include the cost of brine disposal. On the coastal plain of Los Angeles County, brine disposal is not a problem since brines can be returned to the sewerage system and ultimately reach the Pacific Ocean. For inland plants this problem would be of such major proportions as to essentially eliminate consideration of demineralization of reclaimed wastewaters in such locations. Throughout the range of TDS considered, the ion exchange process remains the most economical and amounts to about 20¢ per thousand gallons for a 33% TDS reduction for an influent TDS of 1,000 mg/l. The ion exchange cost curves have therefore been the basis for the generalized cost curves presented in Figures 14 and 15.

Figure 14 shows the estimated cost to obtain a particular product water TDS for a range of influent TDS values up to 1,500 mg/l. Figure 15 shows the estimated cost of achieving a selected amount of TDS removal for the same influent TDS range. These costs include pretreatment by carbon adsorption and exclude the cost of brine disposal. The horizontal plots in Figure 15 illustrate the nature of the ion exchange process since the cost of the process is essentially proportional to the ions removed or exchanged, regardless of the influent TDS. All of the previous estimates have been based on the cost of a 10 mgd demineralization facility. An estimate for a 10 mgd facility is obviously quite an extrapolation from the 2,000 to 15,000 gallons per day pilot plant units which have actually been tested. In Figure 16 the costs of achieving several different product water TDS concentrations with an influent TDS of 1,000 mg/l have been estimated for plant sizes from 0.5 to 10 mgd based on the ion exchange process. It appears that extremely small plant sizes will remain uneconomical but that there may be possibilities for minor improvements in cost for plants larger than 10 mgd. A recent article in the Journal of the American Water Works Association by William R. Gianelli (April, 1970) includes an estimate of 37¢ per thousand gallons for desalting sea water in a 50 mgd plant combined with a large nuclear power generating facility. With no knowledge of what assumptions were actually used in the estimate, it is of interest to note that it is quite comparable to the cost of reducing the TDS in 10 mgd of secondary effluent from 1,000 mg/l to about 500 mg/l obtained from Figure 16.

Summary

The strong acid-weak base ion exchange process piloted at Pomona has proven to be more economical and reliable than the reverse osmosis or electrodialysis systems with which it has been compared. Practical reductions in dissolved solids can be obtained for costs of 20 to 30 cents per thousand gallons where brines can be discharged to the ocean. These costs should be confirmed by construction of a larger pilot facility but indicate that reclamation, demineralization and reuse of wastewater should be considered prior to demineralization of seawater and carefully compared with the estimated costs of future imported supplies.

Credits

The research work reported upon in this paper was jointly financed by the Federal Water Quality Administration and the Sanitation Districts of Los Angeles County. The following project engineers from both agencies have operated the pilot units, collected and analyzed data, and overcame numerous problems in developing the information upon which this paper is based:

Project Engineers	Process
Max Augustus	Reverse Osmosis
Robert Miele	Reverse Osmosis
James Gratteau	Reverse Osmosis
John English	Electrodialysis
John Burgeson	Electrodialysis
Harold Takenaka	Electrodialysis
Hans Krock	Ion Exchange
Ching-lin Chen	Ion Exchange

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PHOTOGRAPH NOT INCLUDED

FIG. 1: POMONA RESEARCH & DEVELOPMENT FACILITY

PHOTOGRAPH NOT INCLUDED

FIG. 2: REVERSE OSMOSIS PILOT PLANTS AT POMONA

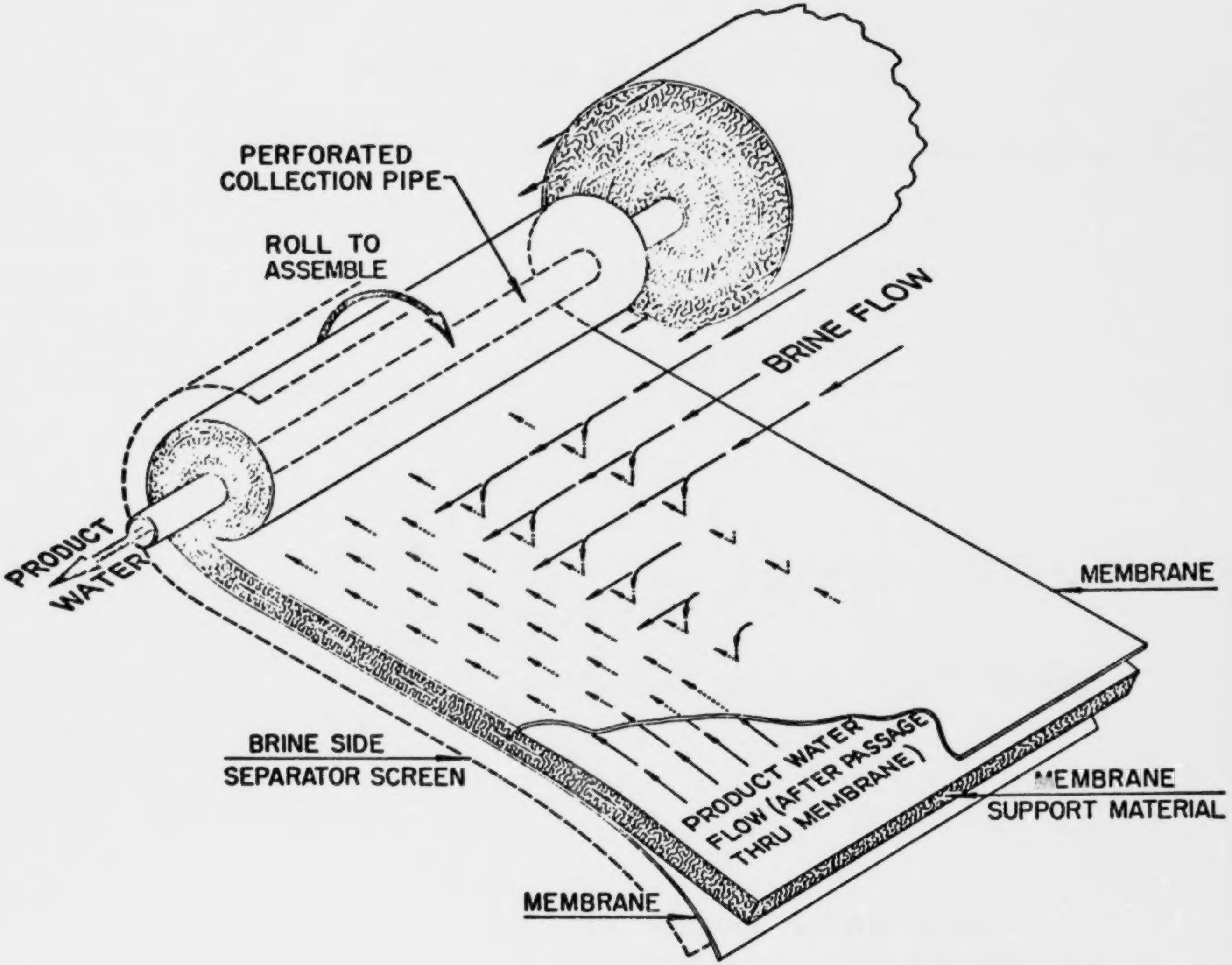
PERFORATED COLLECTION PIPE
ROLL TO ASSEMBLE
BRINE FLOW
PRODUCT WATER
MEMBRANE
BRINE SIDE SEPARATOR SCREEN
PRODUCT WATER FLOW (AFTER PASSAGE THRU MEMBRANE)
MEMBRANE SUPPORT MATERIAL
MEMBRANE

FIGURE 3: SCHEMATIC OF SPIRAL WOUND REVERSE OSMOSIS MODULE

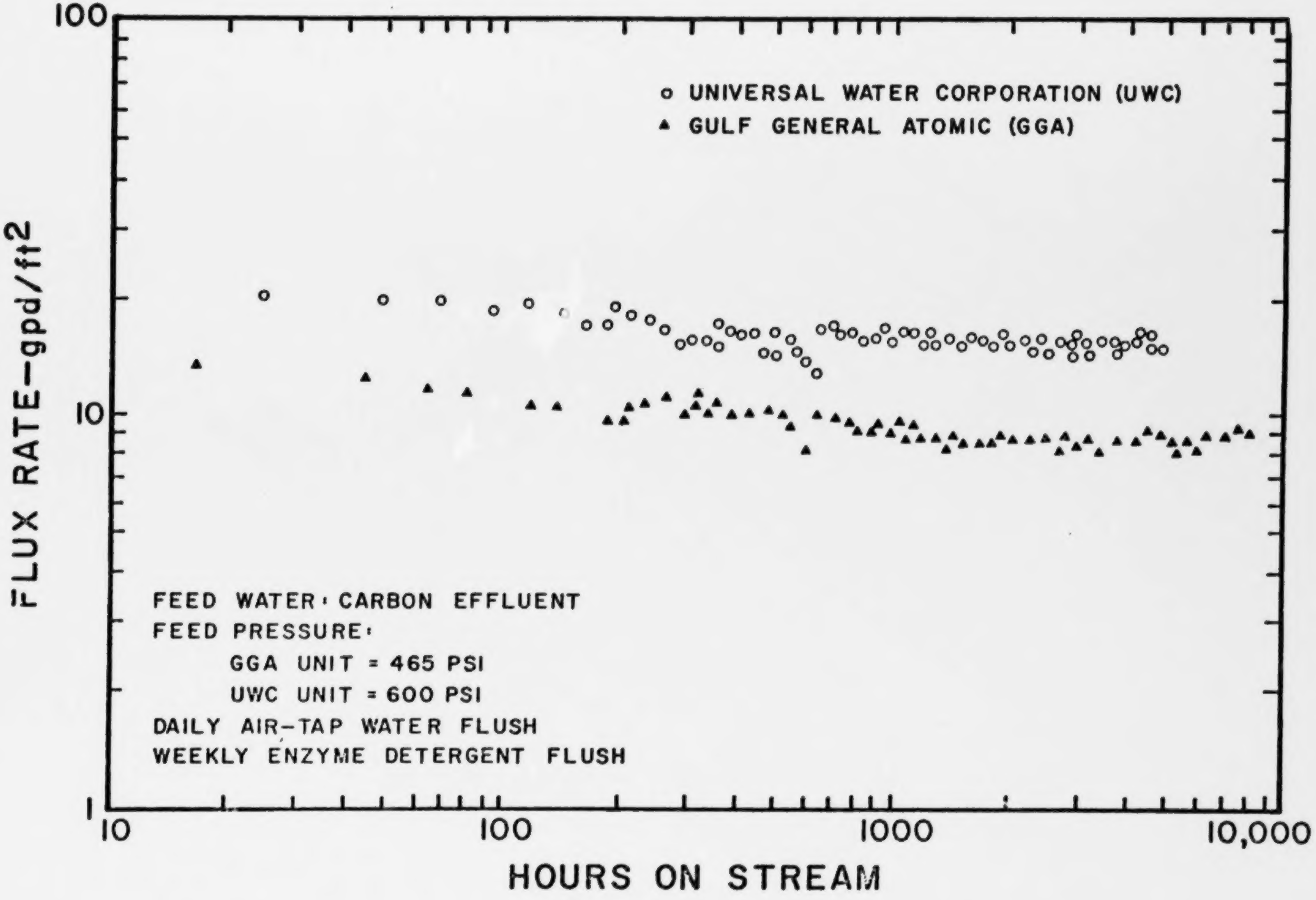

FIGURE 4: PERFORMANCE OF REVERSE OSMOSIS UNITS

PHOTOGRAPH NOT INCLUDED

FIG. 5: ELECTRODIALYSIS PILOT PLANT AT POMONA

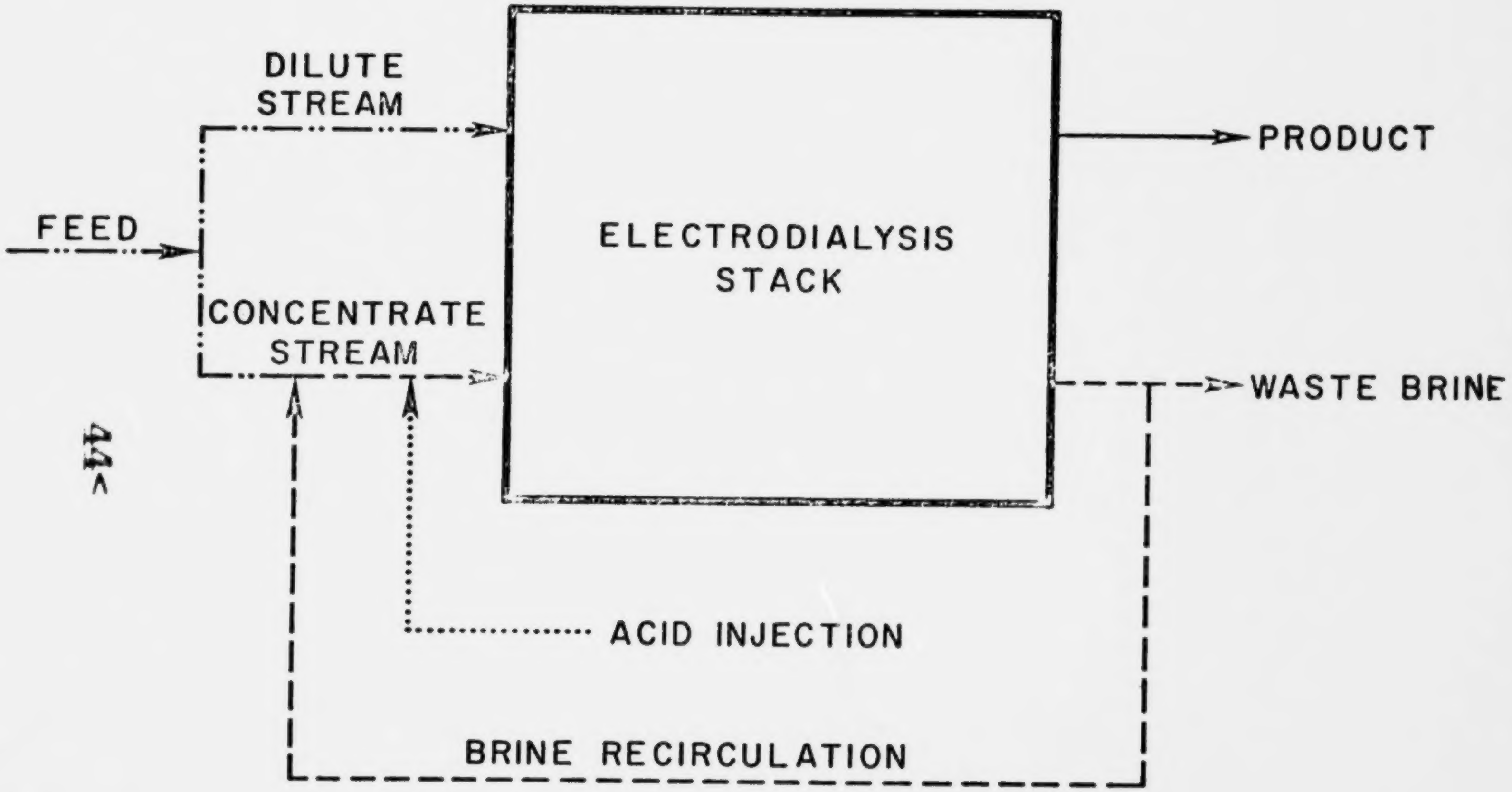
DILUTE STREAM
FEED
CONCENTRATE STREAM
ELECTRODIALYSIS STACK
PRODUCT
WASTE BRINE
ACID INJECTION
BRINE RECIRCULATION

FIGURE 6: SCHEMATIC FLOW DIAGRAM OF ELECTRODIALYSIS PILOT PLANT

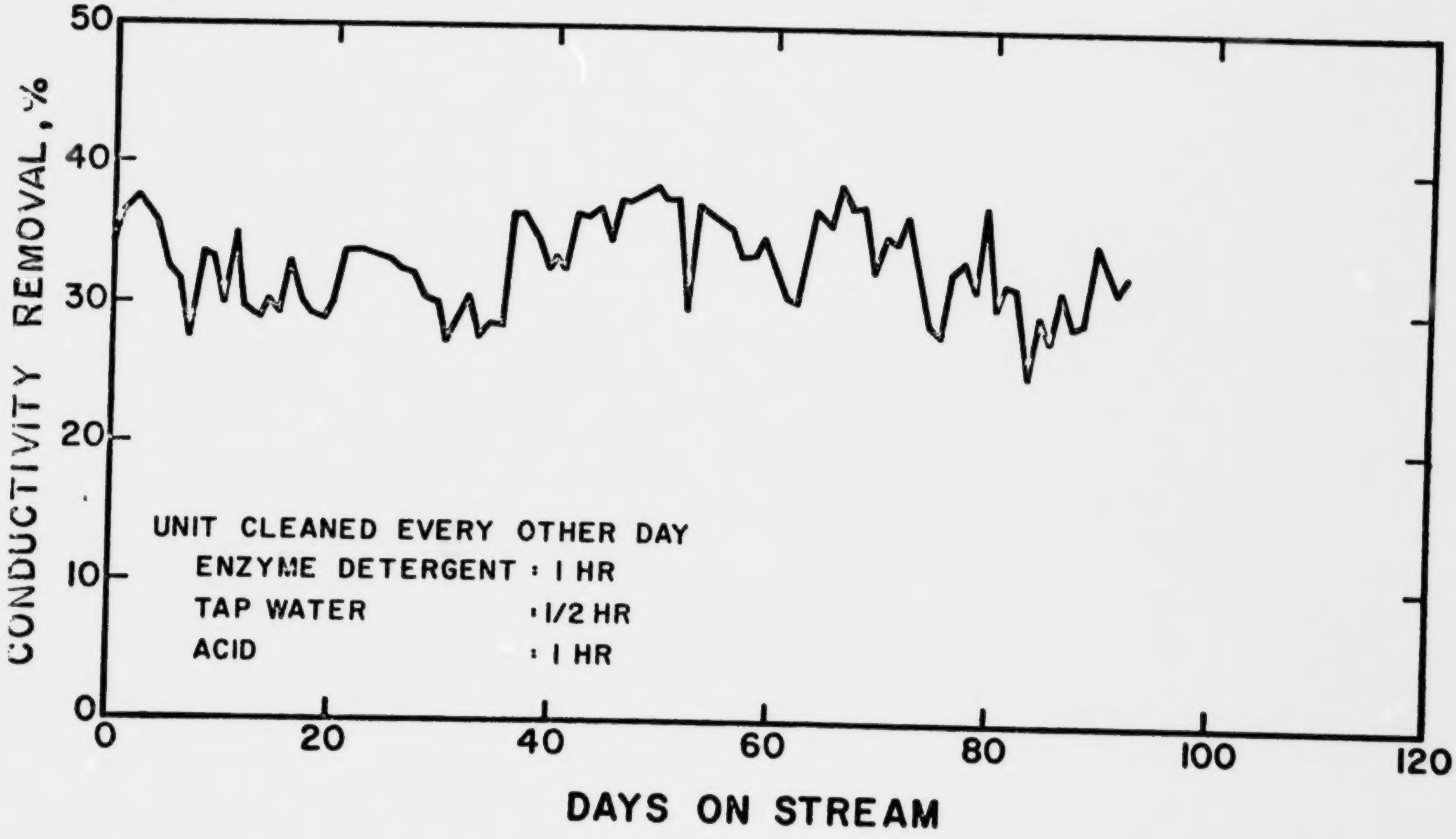

FIGURE 7: PERFORMANCE OF ELECTRODIALYSIS UNIT

PHOTOGRAPH NOT INCLUDED

FIG. 8: ION EXCHANGE PILOT PLANT AT POMONA

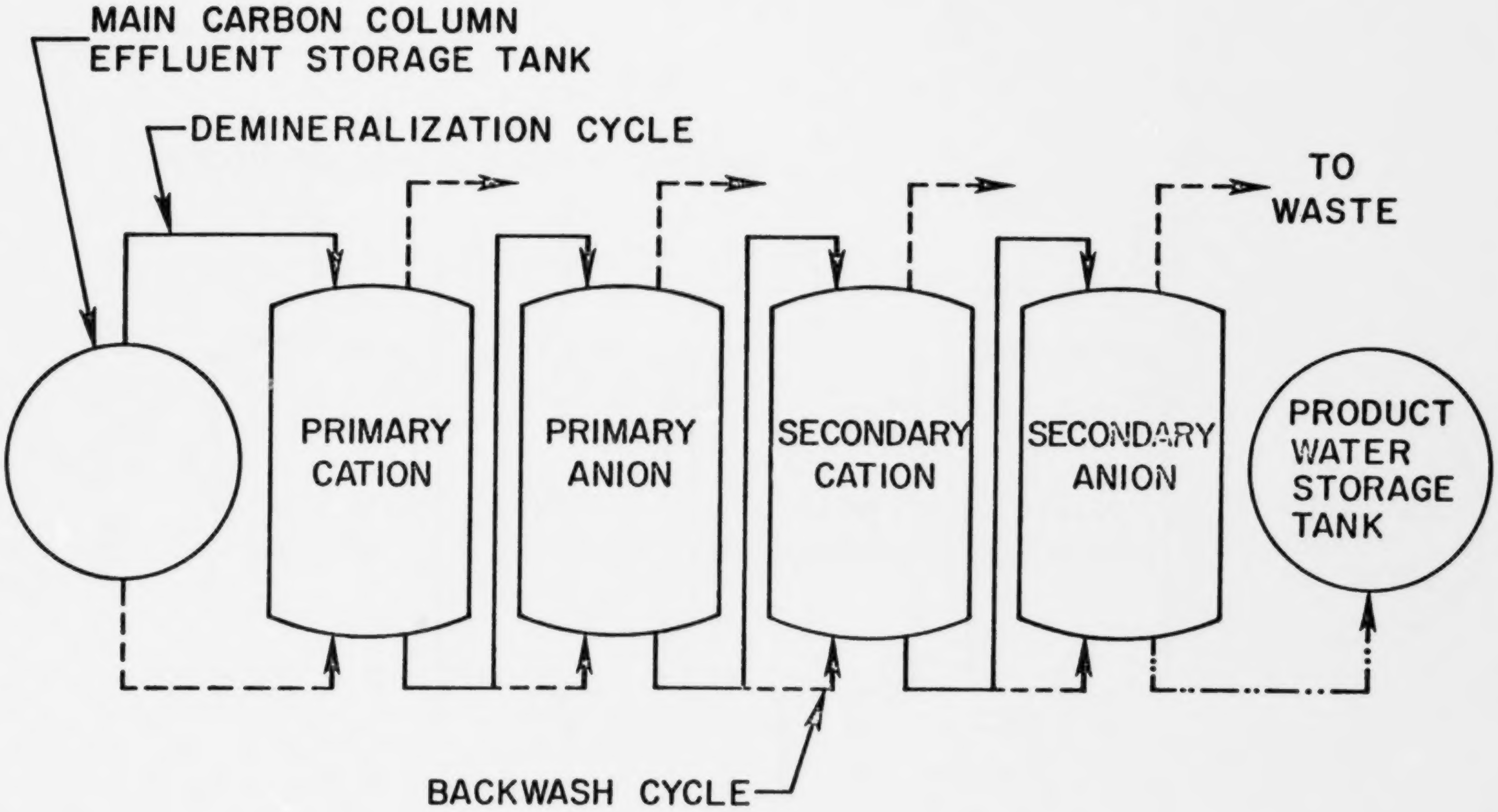
MAIN CARBON COLUMN
EFFLUENT STORAGE TANK
DEMINERALIZATION CYCLE
TO
WASTE
PRIMARY
CATION
PRIMARY
ANION
SECONDARY
CATION
SECONDARY
ANION
PRODUCT
WATER
STORAGE
TANK
BACKWASH CYCLE

FIGURE 9: SCHEMATIC FLOW DIAGRAM OF ION EXCHANGE UNIT

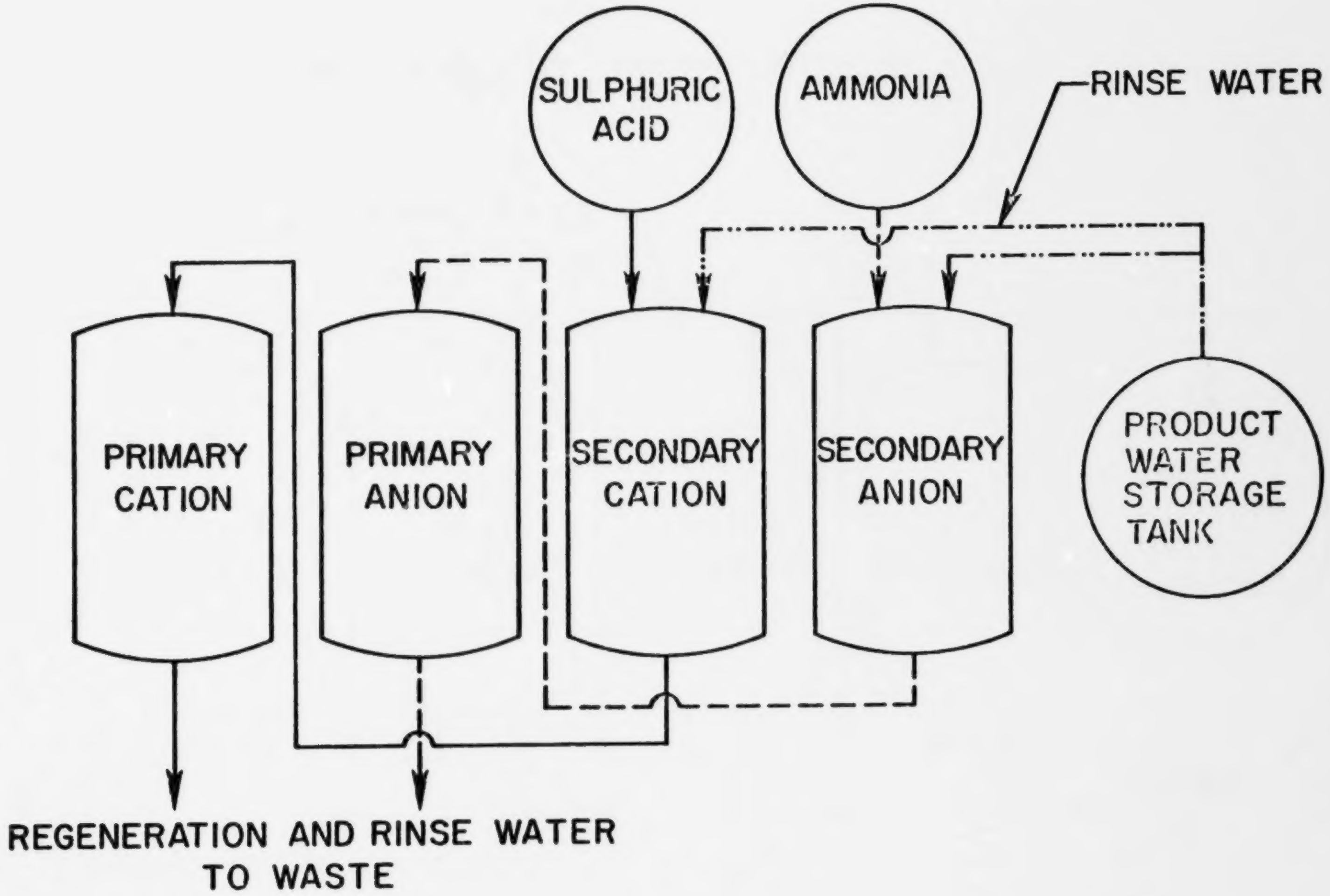
SULPHURIC ACID
AMMONIA
RINSE WATER
PRIMARY CATION
PRIMARY ANION
SECONDARY CATION
SECONDARY ANION
PRODUCT WATER STORAGE TANK
REGENERATION AND RINSE WATER TO WASTE

FIGURE 10: SCHEMATIC OF ION EXCHANGE REGENERATION PROCESS

GPO 621-571-4

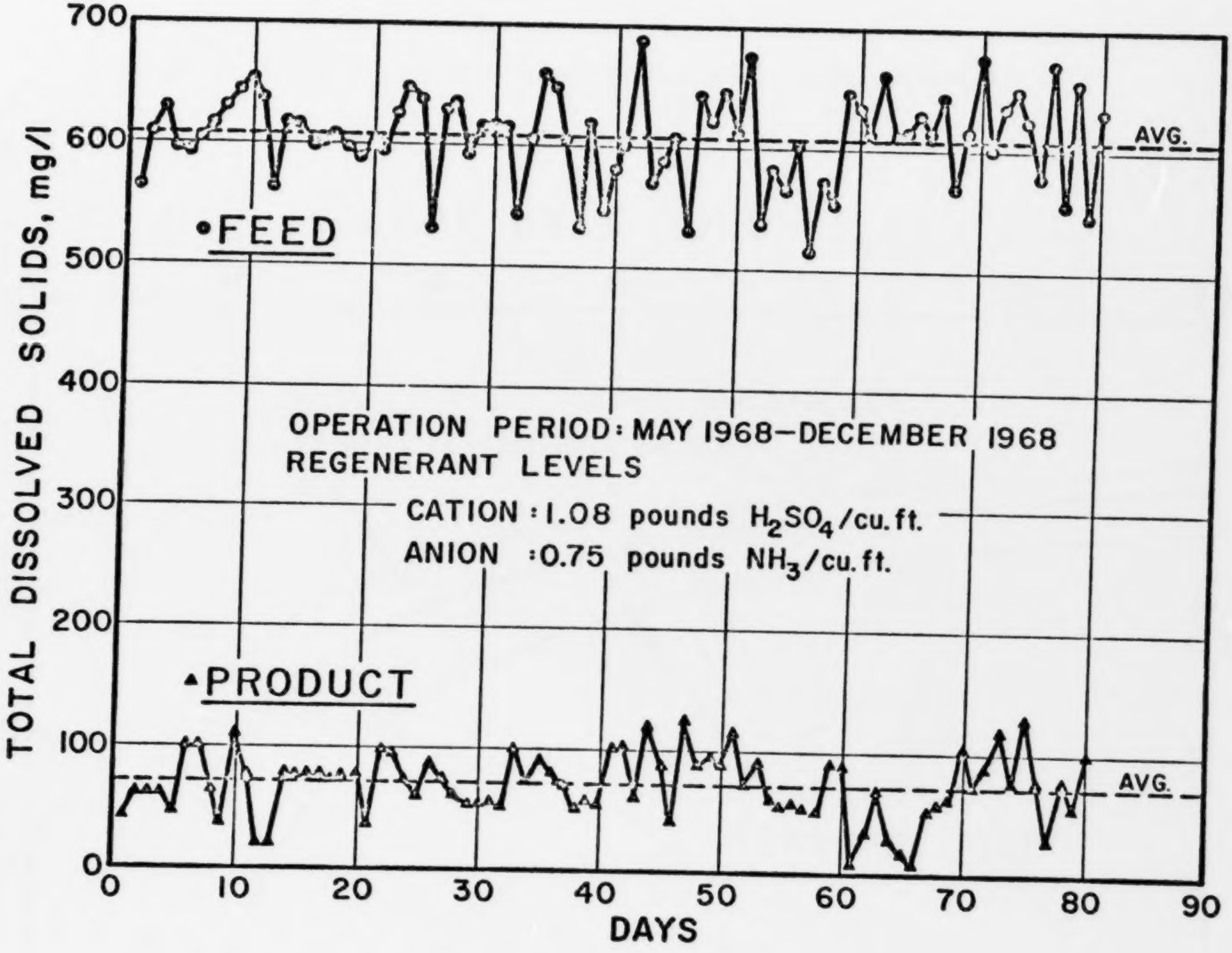

FIGURE 11: PERFORMANCE OF ION EXCHANGE UNIT

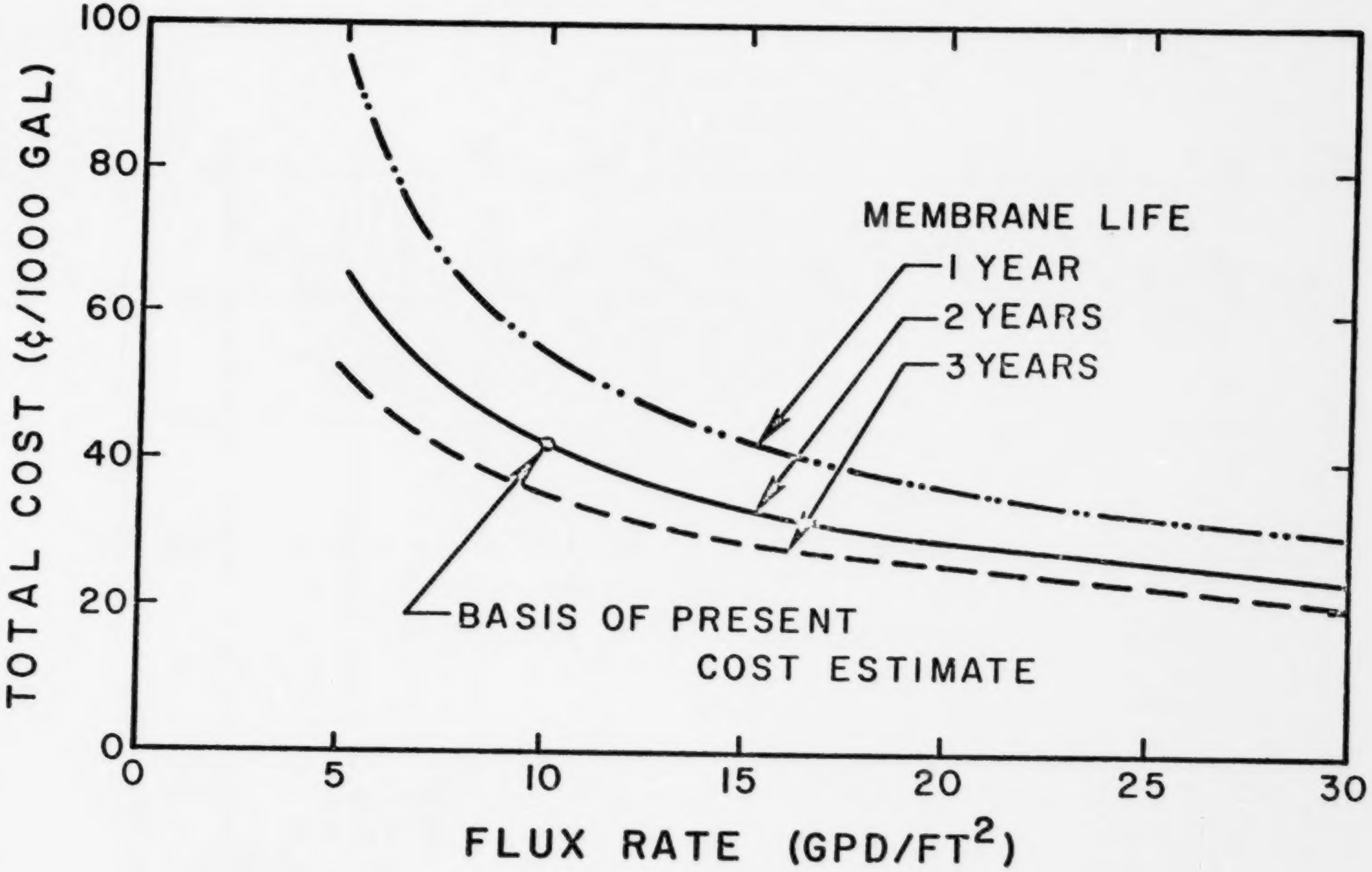

FIGURE 12: EFFECT OF FLUX RATE AND MEMBRANE LIFE ON COST OF REVERSE OSMOSIS

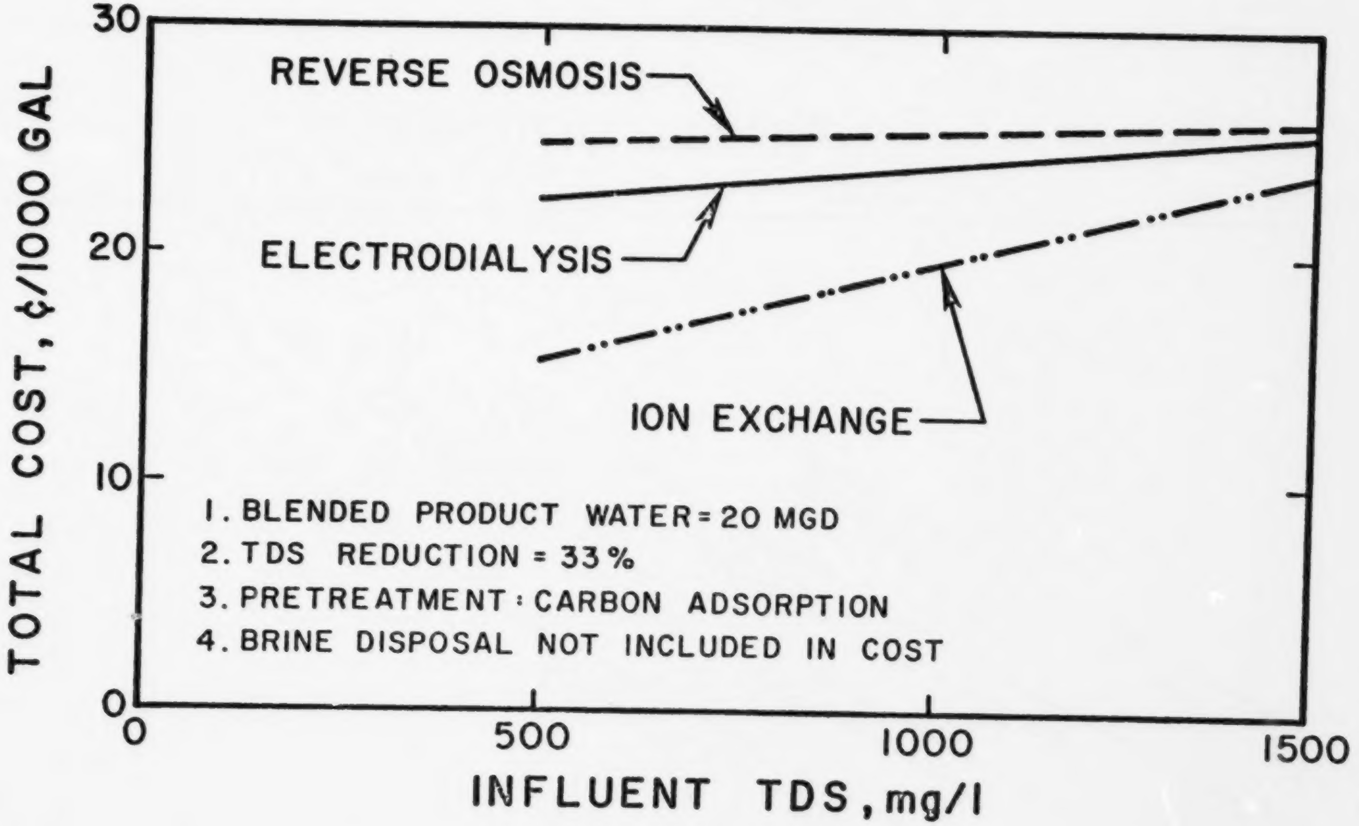

FIGURE 13: COMPARISON OF DEMINERALIZATION COSTS

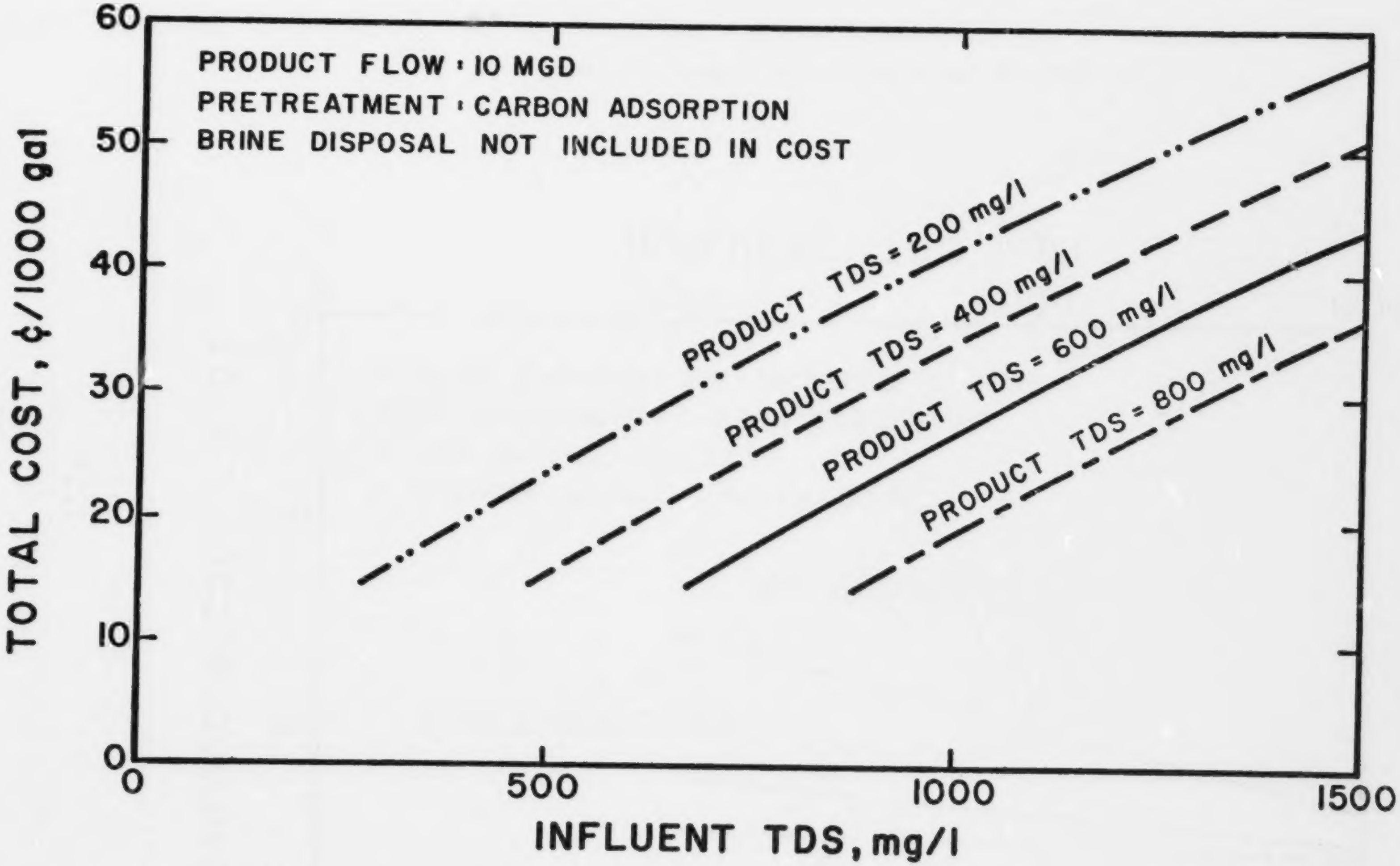

FIGURE 14: COST TO OBTAIN DESIRED PRODUCT TDS

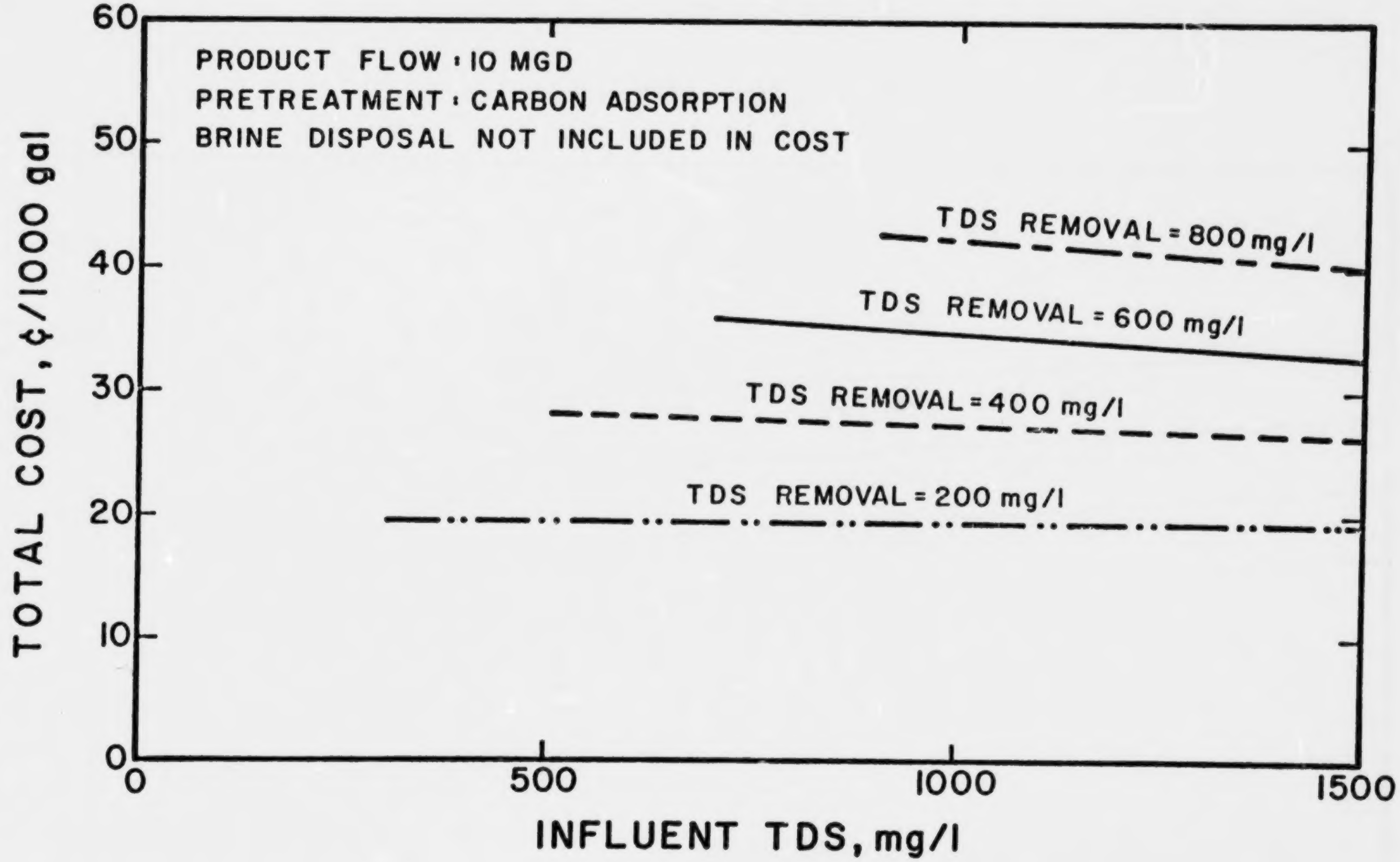

FIGURE 15: COST TO ACHIEVE DESIRED TDS REMOVAL

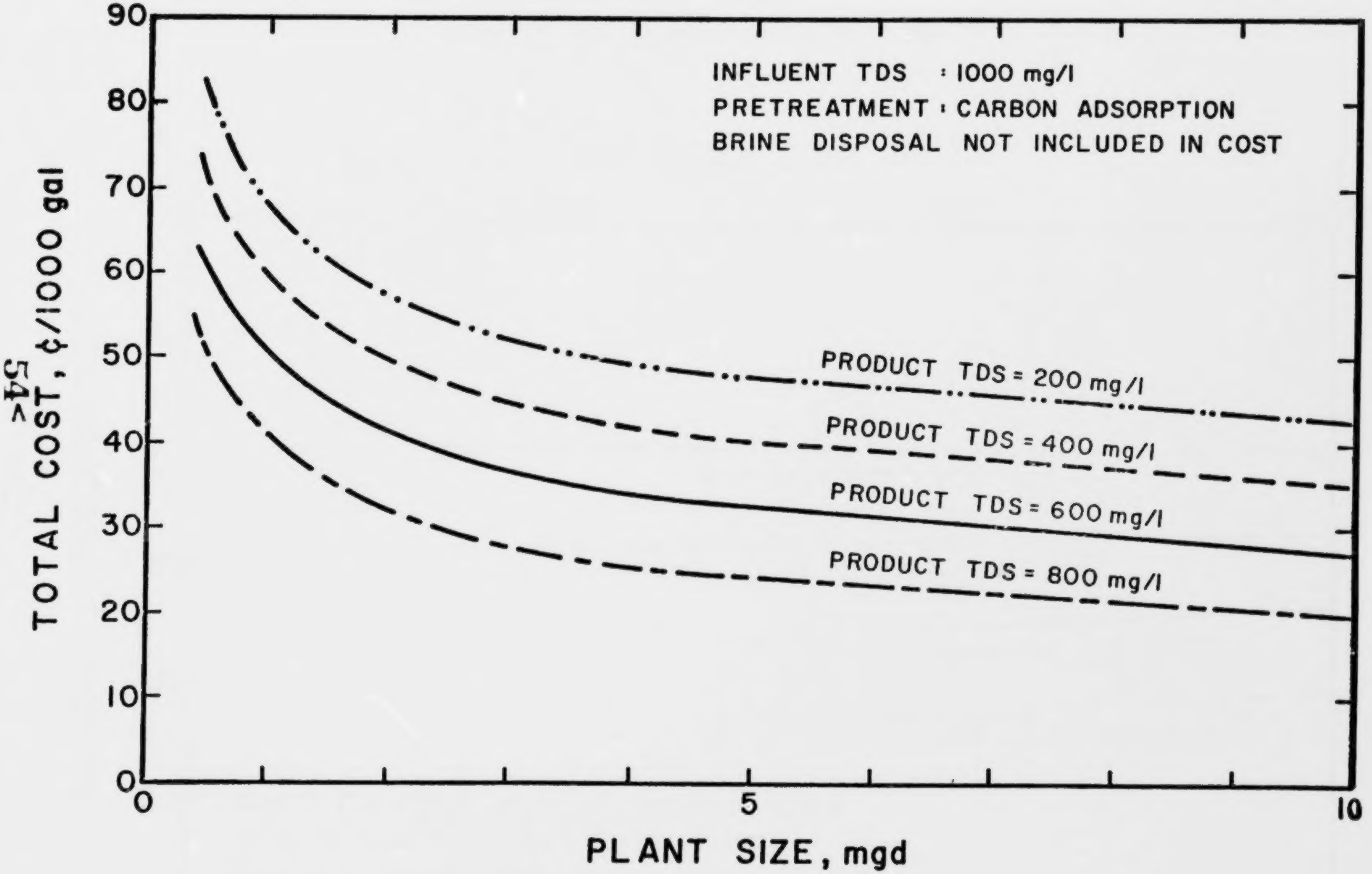

FIGURE 16: EFFECT OF PLANT SIZE UPON COST

PARAMETER		INFLUENT	EFFLUENT
SUSPENDED SOLIDS	mg/l	9	0.6
COD	mg/l	43	10
DISSOLVED COD	mg/l	30	8
TOC	mg/l	12	3
NITRATE, as N	mg/l	8.1	6.6
TURBIDITY	JTU	8.2	1.2
COLOR		28	3
ODOR		12	1
CCE	mg/l	—	0.026

TABLE 1: AVERAGE QUALITY DATA FOR CARBON ADSORPTION PILOT PLANT

TEST (mg/l)	INFLUENT	EFFLUENT	% REMOVAL
COD	8.7	1.0	88.5
NH_3-N	10.1	1.1	89.2
NO_3-N	4.9	2.4	51.0
PO_4-P	10.9	0.2	98.4
TDS	750	59	92.1

NOTES

1. 15,000 GPD PILOT PLANT
2. INFLUENT IS CARBON EFFLUENT
3. AVERAGES BASED ON 40 SAMPLES (JUNE-SEPT. 1969)
4. ANALYSES RUN ON GRAB SAMPLES
5. AVERAGE WATER RECOVERY = 75 %

TABLE 2: WATER QUALITY CHARACTERISTICS OF REVERSE OSMOSIS PROCESS

TEST (mg/l)	INFLUENT	EFFLUENT	% REMOVAL
COD	9.4	8.0	14.9
NH_3-N	9.0	5.1	43.4
NO_3-N	6.2	3.1	50.0
PO_4-P	10.1	7.8	22.8
TDS	705	465	34.0

NOTES

1. 15,000 GPD PILOT PLANT
2. INFLUENT IS CARBON EFFLUENT
3. AVERAGES BASED ON 10 SAMPLES (JUNE-AUG 1969)
4. ANALYSES RUN ON 24-HOUR COMPOSITE SAMPLES
5. AVERAGE WATER RECOVERY =83%

TABLE 3: WATER QUALITY CHARACTERISTICS OF ELECTRODIALYSIS PROCESS

TEST (mg/l)	INFLUENT	EFFLUENT	% REMOVAL
COD	10.0	3.7	63.0
NH_3-N	15.6	3.8	75.6
NO_3-N	2.9	0.4	86.2
PO_4-P	8.8	0.1	98.8
TDS	611	81	86.7

NOTES

1. 5000 GPD PILOT PLANT
2. INFLUENT IS CARBON EFFLUENT
3. AVERAGES BASED ON 20 SAMPLES (MAY-DEC 1968)
4. ANALYSES RUN ON 16-HOUR COMPOSITE SAMPLES
5. AVERAGE WATER RECOVERY = 87%

TABLE 4: WATER QUALITY CHARACTERISTICS OF ION EXCHANGE PROCESS

AMORTIZATION OF CAPITAL		¢/1000 GALLONS
$ 4,020,000 ; 20 YEARS @ 5%		8.8
OPERATION AND MAINTENANCE		
CHEMICALS	7.3	
MEMBRANE REPLACEMENT	14.4	
MAINTENANCE MATERIALS	2.6	
POWER	7.0	
LABOR	1.5	
		32.8
TOTAL		41.6

ASSUMPTIONS

1. INFLUENT TDS=700 mg/l ; EFFLUENT TDS=100 mg/l
2. MEMBRANES : FLUX RATE = 10 GPD/FT^2 ; LIFE = 2 YEARS
3. WATER RECOVERY = 85%

TABLE 5: COST ESTIMATE FOR 10 MGD REVERSE OSMOSIS PLANT

AMORTIZATION OF CAPITAL		¢/1000 GALLONS
$1,830,000; 20 YEARS @ 5%		4.0
OPERATION AND MAINTENANCE		
CHEMICALS	4.4	
MEMBRANE REPLACEMENT	3.3	
MAINTENANCE MATERIALS	0.9	
POWER	1.9	
LABOR	2.5	
		13.0
TOTAL		17.0

ASSUMPTIONS

1. INFLUENT TDS = 700 mg/l; EFFLUENT TDS = 450 mg/l
2. MEMBRANE LIFE = 5 YEARS
3. WATER RECOVERY = 90%

TABLE 6: COST ESTIMATE FOR 10 MGD ELECTRODIALYSIS PLANT

AMORTIZATION OF CAPITAL		¢/1000 GALLONS
$1,370,000 ; 20 YEARS @ 5%		3.0
OPERATION AND MAINTENANCE		
REGENERATION		
CATION (SULFURIC ACID)	10.0	
ANION (AMMONIA)	4.2	
RESIN REPLACEMENT	2.0	
MAINTENANCE MATERIALS	0.6	
POWER	0.8	
LABOR	1.4	
		19.0
TOTAL		22.0

ASSUMPTIONS

1. INFLUENT TDS = 700 mg/l ; EFFLUENT TDS = 75 mg/l
2. REGENERATION EFFICIENCY : CATION = 85% ; ANION = 90%
3. RESIN REPLACEMENT : CATION = 10% AND ANION = 20% PER YEAR
4. WATER RECOVERY = 87%

TABLE 7: COST ESTIMATE FOR 10 MGD ION EXCHANGE PLANT

Franklin D. Dryden
and Gerald Stern

RENOVATED WASTE WATER CREATES RECREATIONAL LAKE

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Renovated Waste Water Creates Recreational Lake

In pilot studies aimed at converting a borrow pit to an aquatic park, the major hurdle proved to be the reduction of phosphorus to concentrations that limit algal growth

Lancaster, Calif., is a rural community of about 35,000 on the southern edge of the Mojave Desert about 80 miles northeast of Los Angeles. For more than 20 years the waste water of the area has been given primary treatment followed by biological stabilization in oxidation ponds. Presently, 3 million gallons per day are treated and then lost to the atmosphere by evaporation. With annual rainfall of only 2 to 3 inches, the area depends on underground sources for both irrigation and domestic water supplies. Supplemental water will eventually be imported from Northern California. In the interim, ground water must be conserved for domestic, industrial, and irrigational purposes. In recognizing the community's need for water-based recreational facilities, the Los Angeles County Engineer proposed converting an existing borrow pit to an aquatic park featuring three lakes with a combined surface area of 19 acres.

The Sanitation Districts of Los Angeles County joined with the Advanced Waste Treatment Branch of the Federal Water Pollution Control Administration, Department of the Interior (formerly the U.S. Public Health Service), in a research effort to investigate the best possible methods of renovating the oxidation pond water to a point where it would be suitable for use in the lakes. Since oxidation ponds are used in many arid areas, research groups hoped that techniques developed in Lancaster would be applicable to other localities.

The water in the oxidation ponds is unsuitable for use in the lakes without further treatment. The green algae which abound in the oxidation ponds supply oxygen to the bacteria which decompose the organic matter in the waste water. However, since algae receive their energy from light they photosynthesize and produce oxygen only during daylight hours. The water becomes supersaturated with dissolved oxygen with maximum concentrations as high as 40 milligrams per liter in Lancaster. The bacteria, however, utilize oxygen continuously and can reduce the dissolved oxygen of the water to as low as 1 mg./l. or less at night. Fish cannot survive in such an extreme environment, a fact confirmed by previous attempts to raise carp in the oxidation ponds.

In addition to sunlight, algae require carbon, nitrogen, phosphorus, and certain trace elements to live. The carbon is obtained from carbon dioxide and bicarbonate in the water, raising the pH to between 8.5 and 9.5, which is a difficult range for fish. Therefore, both the bacterial and algal populations in the oxidation pond water would have to be drastically reduced for the water to be suitable for use in recreational lakes. The bacterial population could be reduced by reducing its food supply—the organic matter and algae in the water. However, a new cycle of algae growth, organic matter production, and bacterial decomposition is sure to start unless one of the essential algal nutrients is also removed. The only essential nutrient that could readily be reduced to growth-limiting concentrations is phosphorus, and its removal has become a major objective of the study.

Research objectives

The immediate study objective was to determine and pilot the most economical method of renovating the waste water for use in the recreational lakes. However, the lack of recognized limiting quality criteria for recreational lake water required that arbitrary criteria be established as contingent objectives in evaluating processes. But these criteria were flexible and could be revised as the study revealed cost data, process capabilities, and what may be the special behavior of renovated water.

Reviews of studies by Sawyer (1947) and Malhatra (1964), consultation with algologists at the Robert A. Taft Sanitary Engineering Center, and knowledge of the recent experience of the Santee County Water District with recreational lakes supplied by soil-filtered waste water near San Diego (1966) provided the basis for the following initial quality objectives:

- Turbidity of less than 10 Jackson turbidity units (JTU).
- Phosphate concentration of less than 2 mg./l. as PO_4.
- pH of less than 8.0.
- Low concentration of chemical oxygen demand (COD) and biological oxygen demand (BOD).
- Algae counts less than 10,000 per milliliter.
- Coliform and virus determinations to meet requirements of the County Health Officer.

Subsequent experience has resulted in the revision of the phosphate objective from 2 mg./l. to 0.5 mg./l. or less. The other criteria are still considered adequate. Product water, or water produced by the selected process, is generally of better quality than would be indicated by these objectives. The turbidity objective of 10 JTU, for example, is double the average turbidity of the product water. The final turbidity is a consequence of the chemical dosage required for phosphate removal. The low turbidity makes the water clear and sparkling, which will increase its public acceptance as it enters the lake.

Oxidation pond facilities

The District 14 Water Renovation Plant features primary sedimentation followed by 240 acres of oxidation ponds. (Figure 1 shows the project site, facilities, and oxidation pond flow pattern.) Each of the eight ponds has a surface area of about 30 acres and 5- to 7-

FEATURE

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Gerald Stern, Southern California Field Station, FWPCA, Garden Grove, Calif. 92640

foot water depth. Only the first four ponds are in the treatment series; the last four are used for winter storage when evaporation rates are low and are dried out in the summer to inhibit midge propagation.

Primary effluent is divided equally between Ponds 1 and 2 and dispersed through 400-foot diffusers to avoid overloading a portion of the pond. Pond 1 discharges into Pond 2; the combined flow then passes sequentially through Ponds 4, 3, and 5 as shown. The flow pattern can be varied as desired. The average residence time to the discharge of Pond 3 is between 45 and 60 days. Pond 3 is the designated source for water to be used in the recreational lakes and was the source for all water treatment studies conducted in this project.

The project was conducted in two phases:

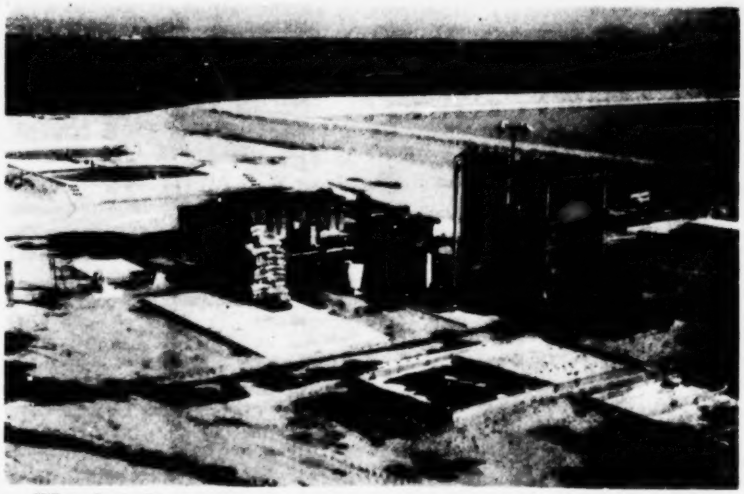
- Determining the most effective and economical process for renovation of the waste water (Phase I).
- Operating the chosen process and evaluating the product water in simulated recreational lakes (Phase II)

Phase I studies

To accomplish the desired objectives in minimum time, three research steps that would normally be conducted successively were performed concurrently. They are:

- Collection of basic data on the character of the water to be treated.
- Laboratory jar test evaluations of various coagulants to determine the most feasible chemical treatment and the optimum conditions for removing algae, other suspended material, and nutrients.
- Pilot-plant process evaluation of promising processes to develop cost and performance data in order to select preferred processes for renovating the waste water.

Analyses of the plant primary effluent



Pilot plant. *Six processes for water reuse were evaluated at Lancaster, Calif.*

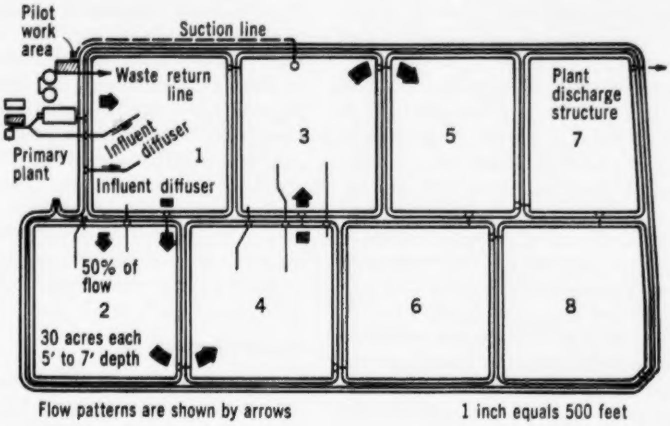

Pilot work area
Suction line
Waste return line
Influent diffuser
Primary plant
1
Influent diffuser
3
5
Plant discharge structure
7
50% of flow
2
30 acres each
5' to 7' depth
4
6
8
Flow patterns are shown by arrows
1 inch equals 500 feet

Figure 1

Oxidation ponds and study area

64<

and of the oxidation pond water began in October 1963. Grab samples were collected from the pond and analyzed for selected chemicals and algae genera and numbers. An automatic sampler was installed on the primary effluent in July 1964, to obtain more representative data. The average characteristics of the primary effluent are:

pH	7.2
Total alkalinity	258 mg./l.
Phosphate	45 mg./l.
Ammonia-nitrogen	27 mg./l.
COD	392 mg./l.
Suspended solids	90 mg./l.
Dissolved solids	553 mg./l.
Organic nitrogen	11 mg./l.
Nitrate-nitrogen	0.1 mg./l.
Nitrite-nitrogen	<0.01 mg./l.
Hardness	77 mg./l.
Calcium	32 mg./l.
Magnesium	3 mg./l.
Sulfate	70 mg./l.

Pondwide sampling established that the long detention times within the ponds and frequent strong winds resulted in excellent mixing and averaging of most chemical constituents at any given time.

In the course of the study we observed that, to achieve equivalent results, greater chemical doses were required during the winter and spring period of high total alkalinity, phosphate, and ammonia than during the summer and fall period when lower values prevailed. Essentially all chemical changes in Pond 3 were traced to either physical factors such as sunlight intensity, temperature, and wind mixing, or biological factors such as specie composition of the biomass, metabolic rates of the algae and bacteria, and growth of crustaceans.

Ammonia concentration showed a cyclic pattern—less than 3 mg./l. during the summer and fall, increasing to about 20 mg./l. for much of the winter and spring (Figure 2). The seasonal decrease of ammonia by biological synthesis represents most of the nitrogen removal that is accomplished in the oxidation ponds. The low ammonia levels in the summer correlate with the period of rapid metabolism. The high levels in the winter correspond to the period when low sunlight intensity and low temperatures reduce the rate of photosynthesis. Nitrate and nitrite concentrations are generally low throughout the year. However, in the spring of 1966 as temperatures were rising, nitrite values reached 10 mg./l. for about a week before the development of nitrate-producing bacteria, which converted some of the nitrite to nitrate. Denitrification may also have occurred; nitrate values never exceeded 2 mg./l., while the nitrite dropped almost to zero. Organic nitrogen has been observed as high as 20 mg./l., coinciding with high algal counts.

Ammonia variations are also accompanied by similar changes in total alkalinity, which averages 260 mg./l. in the winter and spring and 220 mg./l. in the summer and fall. The higher alkalinity and ammonia concentrations, which tend to buffer pH changes, and the lower temperature, which reduces chemical reaction rates, require that additional coagulants be used for equivalent treatment.

Total phosphates

The total phosphate concentration is also higher during the winter and spring than in the summer and fall period. Biological synthesis, sedimentation of algae, and possibly calcium phosphate precipitation at the higher pH values of the summer and fall all contribute to lowering total phosphate levels. The low total phosphate concentrations permit low coagulant dosages for equivalent residuals.

Other periodic variations between the colder and warmer seasons which are important to the tertiary process include:

- Temperature, which varies from 6° to 30° C.
- pH, which ranges from 7.5 to 9.6.
- Chemical oxygen demand, which decreases from 300 to approximately 200 mg./l.
- Biological oxygen demand, which decreases from 50 to 40 mg./l.
- Suspended solids, which decrease from 100 to 75 mg./l.
- Dissolved solids, which decrease from 640 to 570 mg./l.
- Turbidity, which ranges from 90 to 170 JTU in the cold season and from 40 to 150 JTU in the warm season.

Algae genera and predators

The oxidation ponds contain about 66 algae genera in varying concentrations (50,000 to 700,000 counts per ml.) throughout the year. *Scenedesmus* is the most common, although *Euglena*, *Ankistrodesmus*, *Phacus*, *Oöcystis*, *Chlorella*, and *Nitzschia* have also been dominant.

The most common crustaceans found at random in the oxidation ponds at different times are *Daphnia* and *Cyclops*. These algae predators contribute to the variation in algae counts and algae genera.

The oxidation ponds are very effective in reducing both bacteria and virus levels. The bacteria reductions, after 45 to 60 days' detention, were more than 99.7% (Figure 3). Viruses were detected in 96% of primary effluent samples, and in only 8% of the Pond 3 samples. Polioviruses I, II, and III, the adeno group, coxsackievirus B-3, echovirus 7, and reovirus have been identified in the primary effluents. In oxidation Pond 3 waste water, one sample contained polio II and three samples contained an unidentified virus.

Laboratory studies

Laboratory studies included jar test evaluation of coagulants. The object was to determine the most effective coagulants and coagulant aids, their required dosages, the best flocculating conditions, and relative costs.

The jar test apparatus consisted of a 10-place stirrer equipped with a variable speed paddle drive. All tests were performed on water from oxidation Pond 3 following the procedures suggested by Cohen (1957).

To facilitate screening a maximum number of samples, the effectiveness was rapidly estimated by a turbidity measurement. Additional chemical analyses were performed on particularly promising samples.

The coagulant combinations tested are too numerous to report. The major coagulants were:

- Aluminum sulfate [$Al_2(SO_4)\cdot 14H_2O$] as a commercial grade alum, with and without pH adjustments and other coagulant aids.
- Calcium hydroxide [$Ca(OH)_2$] as lime, both high-calcium [90% $Ca(OH)_2$] and the less expensive dolomite [58% $Ca(OH)_2$ and 41% $Mg(OH)_2$].
- Ferric sulfate [$Fe_2(SO_4)_3$], 70% pure, 30% inert.
- Dow C-31 cationic polyelectrolyte. (Dow NC1118, UCAR-149, and Nalco 600 cationic polyelectrolytes were also tested. Dow C-31 test results were very

similar to the other three cationic polyelectrolytes.)

- Sulfuric acid (H_2SO_4) at 66° Baume.

Jar test studies were run to determine costs for the primary coagulants (Figure 4). Dolomitic lime was used in the cost analyses even though high-calcium lime was actually used in pilot studies because it was easier to handle. The jar test showed the two products to be equally effective at equivalent dosages. Ferric sulfate was omitted from further primary coagulant studies due to its higher cost.

The majority of the jar tests were conducted with alum, lime, and various cationic polyelectrolytes as coagulants. In addition, sulfuric acid and carbon dioxide were used for pH adjustment in conjunction with alum coagulation.

Coagulant studies

Tests were also run for primary coagulants and coagulant combinations (Figure 5), including the overall effect of levels of chemical constituents in Pond 3, to determine their cost.

Figure 6 shows the effectiveness and costs of coagulant combinations during periods when the difference in chemical requirements is related to changes in the characteristics of Pond 3 water. Lime was less effective than alum in the spring period, although it was competitive with alum in the fall. As for total phosphate removal, alum coagulation was much more successful than lime, in the colder season, for reducing phosphate concentrations (Figure 7). With lime coagulation, a substantial colloidal precipitate remained, which could account for the higher phosphate residual.

The use of cationic polyelectrolytes or acid for pH adjustments materially reduced the quantities of alum required to accomplish turbidity objectives in each seasonal period, although the coagulant cost in the spring was much higher than in the fall. However, cationic polyelectrolytes and sulfuric acid do not remove soluble phosphate. Consequently, alum in combination with acid or polyelectrolytes is not as effective in removing phosphate so the larger dosage of alum alone. Since alum was the most effective coagulant, it received the most attention in pilot work.

Jar test studies were also used to determine optimum conditions for successful alum coagulation of Pond 3

Figure 2

Phosphate and ammonia concentrations in Pond 3

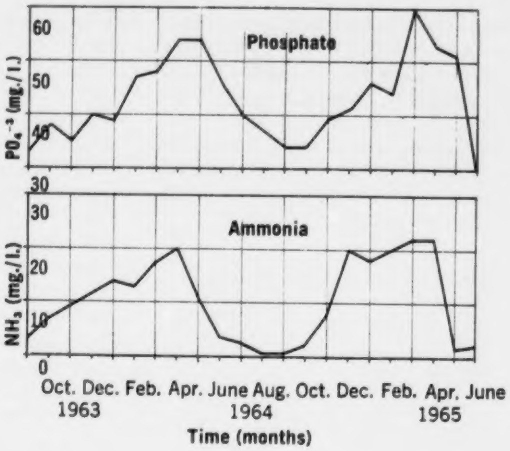


Figure 3

Median bacteria counts through oxidation ponds

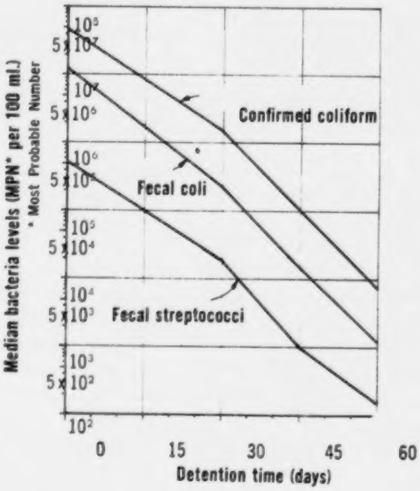


Figure 4

Cost vs. removal efficiency for primary coagulants based on jar tests

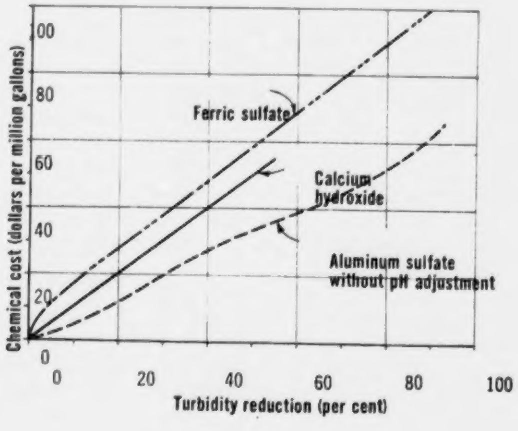


Figure 5

Cost vs. final turbidity for different coagulants based on jar tests

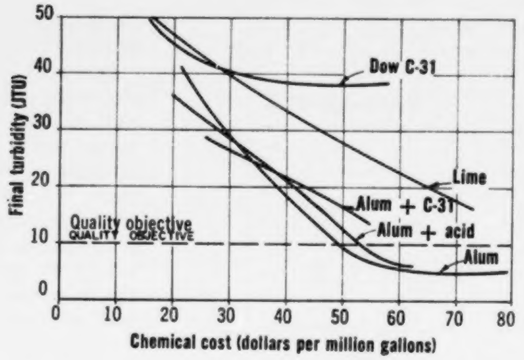
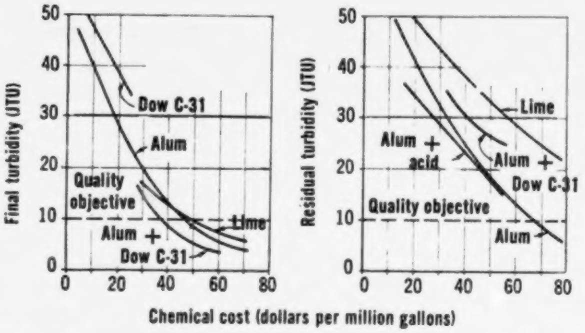


Figure 6

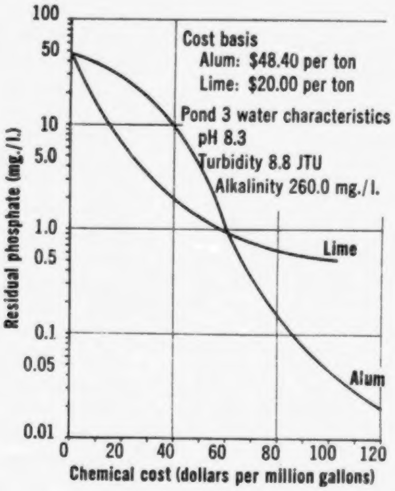
Cost vs. final turbidity for different coagulants during high and low alkalinity and nutrient period, based on jar tests



Low alkalinity and nutrient period	High alkalinity and nutrient period
Alkalinity 220 mg./l.	Alkalinity 260 mg./l.
Phosphate 35 mg./l.	Phosphate 55 mg./l.
Ammonia nitrogen 2 mg./l.	Ammonia nitrogen 20 mg./l.

Figure 7

Residual phosphate for alum and lime coagulation based on jar tests



water for phosphate removal. These studies showed that with or without the addition of acid and base to change the initial pH and alkalinity, phosphate residuals were minimum when the final pH was approximately 6.0 (Figure 8). Increasing the alum dosage from 150 to 300 mg./l. also decreased the minimum phosphate residual. Others have found that the equilibrium constant for alum indicated phosphate can be removed to 0.04 mg./l. and that the pH for optimum removal is close to 6.0.

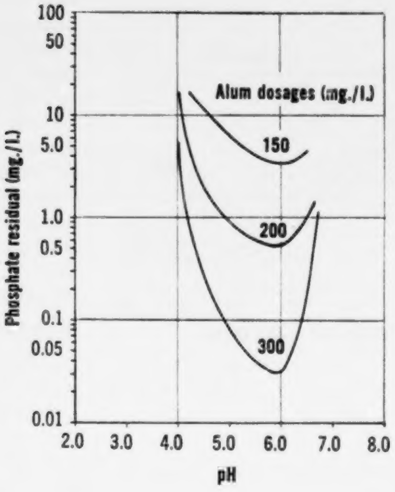
Pilot plant studies

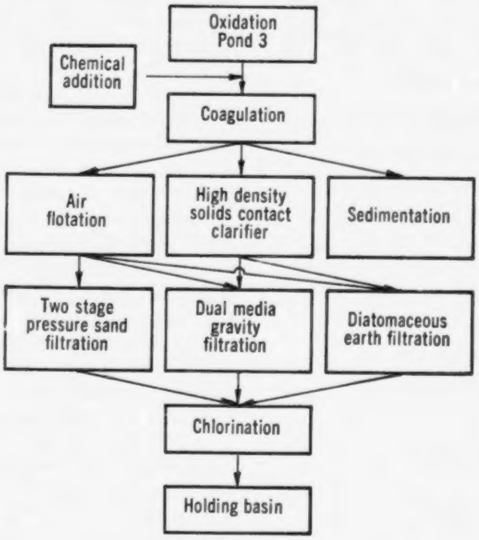
The pilot plant studies were designed to evaluate and compare various treatment processes and to obtain reasonable cost estimates. The pilot area was arranged around a central run of influent and effluent piping with connections for six concurrent processes. Oxidation pond water was delivered to the pilot area from mid-depth on the north side of Pond 3 through a 1600-foot, 6-inch-diameter pipeline, and waste and product water were pumped back into Pond 1. Pilot units ranged in capacity from 1 to 50 gallons per minute (g.p.m.).

Initially, experiments were conducted to remove algae separately from the dissolved nutrients in the hope of producing a marketable algae for use as a poultry feed. A small microstrainer (23-micron screen), a pressure sand filter, and three types of diatomaceous earth filters were tested. These processes failed to produce the desired results. Removals with the microstrainer were totally inadequate and blinding by the bodies of crustacea and other foreign material occurred quite rapidly. The pressure sand filter removed a good percentage of the algae but not enough to be economical. Diatomaceous earth was quite successful in removing the algae, although color and turbidity of the product water were still high. The estimated value of the recovered algae was $25 per million gallons (based on $100 per ton), but the process costs for short filter cycles, separating the algae from the cake, and drying them, far exceeded the product value. More recent evaluations by Dr. William Oswald (University of California, Berke-

Figure 8

Phosphate residual vs. pH based on jar tests




Oxidation Pond 3
Chemical addition
Coagulation
Air flotation
High density solids contact clarifier
Sedimentation
Two stage pressure sand filtration
Dual media gravity filtration
Diatomaceous earth filtration
Chlorination
Holding basin

Pilot plant flow sheet (phase I)

Figure 9

ley) have shown some algae to be a substitute for fish meal, value per ton: $200.

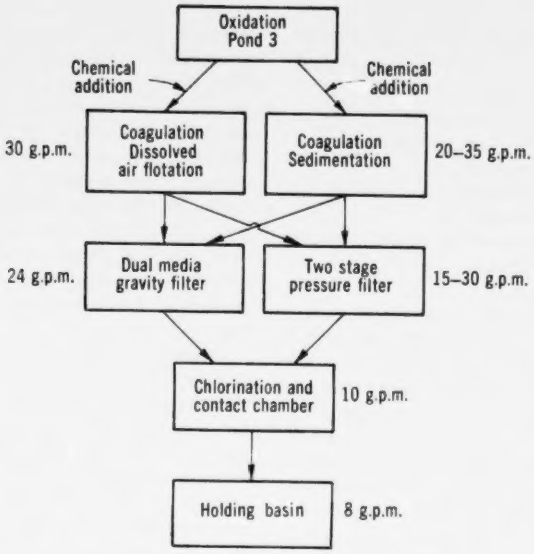
Thereafter, the study concentrated on the best possible processes for clarifying the water. In spite of the potential savings to be gained by completely removing turbidity and dissolved nutrients in a single process, preliminary pilot work indicated that a combination of processes would be required to achieve the desired quality objectives. The first process would materially reduce suspended matter and nutrients by chemical treatment, and the water would be polished by filtration and disinfected by chlorination (Figure 9).

Chemical processes

Initially, the quality objectives for the chemical processes were a turbidity of 20 JTU or less and total phosphate concentration of less than 2 mg./l. Subsequent algal regrowth studies emphasized the need to reduce total phosphate levels to less than 0.5 mg./l. by using larger dosages of alum, which also reduced turbidity levels to the 5- to 10-JTU range. Turbidity was used to compare the relative effectiveness of the pilot processes in screening tests because of the ease and rapidity of measurement. The following types of pilot equipment were tested:

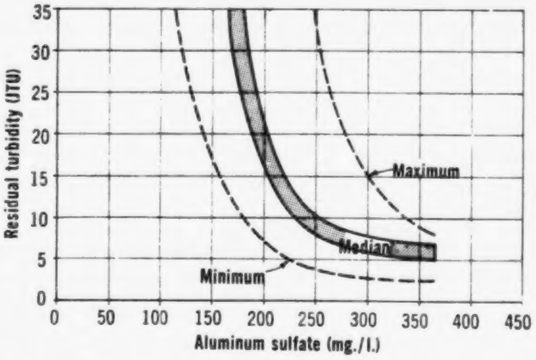
Portable water treatment plant. A completely self-contained trailer-mounted water purification unit of 25-g.p.m. rated capacity loaned to the project by the U.S. Army. It featured an upflow clarifier with a central mixing and flocculating chamber. Results were not satisfactory because of the release of supersaturated dissolved oxygen and other gases which buoyed the floc to the surface.

High density solids contact clarifier. A complete lime-coagulation and recarbonation water treatment unit of 25-g.p.m. capacity with provision for sludge recycle and auxiliary chemical addition. The first of two 10-foot high, 4-foot diameter vertical columns contains rapid mix, coagulation, and flocculation zones near the top with upflow clarifier and sludge collection zones in the bottom. The second column or recarbonator contains inverted trays


Oxidation Pond 3
Chemical addition
Chemical addition
30 g.p.m.
Coagulation Dissolved air flotation
Coagulation Sedimentation
20–35 g.p.m.
24 g.p.m.
Dual media gravity filter
Two stage pressure filter
15–30 g.p.m.
Chlorination and contact chamber
10 g.p.m.
Holding basin
8 g.p.m.

Continuous pilot plant flow sheet (phase II)
Figure 10

Figure 11
Residual turbidity vs. aluminum sulfate prior to filtration (pilot plant)



to adjust pH by adding CO_2 and a chamber for supplementary settling and sludge recycle. The unit did not operate successfully at its rated capacity but was adequate at 10 to 15 g.p.m.

Dissolved air flotation units. Pilot units in two sizes were used in this study:

- A 1-square foot air flotation unit—rated capacity 0.3 to 3.0 g.p.m., air dissolved into auxiliary water stream at 60 to 70 pounds per square inch (p.s.i.), manually skimmed, operated at about 1 g.p.m.
- A 6-foot diameter, 27-square foot × 6-foot deep air flotation unit rated at 50 g.p.m. Coagulated water entered a supersaturated mixing chamber in the center of the 6-foot high tank. Air dissolved into the recycle water is released in small bubbles that lift the floc to the top where it is skimmed into a trough. The clarified water flows to the bottom of the tank and up to and over a circular weir around its periphery. The unit was operated at flow rates between 20 and 50 g.p.m. with most runs at 30 or 40 g.p.m. The recycle stream was varied from 10 to 25 g.p.m. at pressures of 40 and 20 p.s.i., respectively. Air at about 15 cubic feet per hour was introduced by eduction into the suction of the recycle pump.

Sedimentation tank. A 1200-gallon circular tank, 4 feet deep and 7.5 feet in diameter, was baffled to serve for a brief trial of horizontal-flow sedimentation. At 11 g.p.m. there was a 2-hour detention time and an overflow rate of 340 gallons per day (g.p.d.) per square foot. A submerged discharge protected the effluent from floating algae. It was sufficiently effective to justify construction of a larger unit.

The high density solids contact clarifier, air flotation, and sedimentation processes were technically feasible. The sedimentation tank test results were almost identical to those obtained in the air flotation unit in contrast to the very poor results obtained in the truck-mounted upflow clarifier used in the early part of the study. The pilot-plant tests substantiated the jar tests results in that alum remained the most effective chemical coagulant. Thus, both air flotation and horizontal sedimentation with alum as primary coagulant were chosen as the preferred processes to be evaluated in Phase II.

Filtration equipment. Polishing of primary process effluent by filtration provides a safety factor and produces a water of such clarity that its appearance will be pleasing to the public. Both diatomaceous earth filters and sand filters were tested. Diatomaceous earth filters were the following types:

- A low pressure reversible cake diatomaceous earth filter of 0.5 to 3 g.p.m., operated at approximately 1 g.p.m. for pilot tests.
- A truck-mounted 25-square foot pressure filter.
- A wound-wire pressure filter.
- A vertical leaf swimming pool pressure filter.
- A 1-square foot bench-scale pressure filter.
- A 50-square foot tube and shell pressure filter.

Sand filters tested were:

- A two-stage pressure sand filter of 2-foot depth and 2.2-square foot surface area per stage.
- Six 3½-inch-diameter, 0.06-square foot lucite gravity sand and dual media filter columns.
- A 12-square foot gravity sand or dual media (anthrafilt-sand) filter.

Diatomaceous earth filter effluents were consistently better than effluents from the dual media or pressure sand filters. However, diatomaceous earth filter cycles did not exceed 3 hours, resulting in excessive diatomaceous earth usage and high costs (estimated between $80 and $200 per million gallons). The short filter cycles were caused by blinding of the filter by algae and other particulate matter not removed in the previous separation process. The compressibility of the gelatinous floc contributed to the rapid blinding in spite of diatomaceous earth body feed rates of up to 90 mg./l. and influent water turbidities as low as 3 JTU.

Pressure sand and dual media filtration both reduced influent turbidities by 25 to 50% resulting in filter effluents with turbidities of 10 JTU or less when the effluent turbidity from the chemical process was between 10 and 20 JTU. Longer operating cycles and the reuse of the filter media provided a substantial cost advantage for sand and dual media filter over the diatomaceous earth units.

Preferred processes

The combined processes to accomplish the quality objectives at lowest cost were:

- Alum coagulation.
- Either dissolved air flotation or plain sedimentation.
- Either pressure sand or dual media gravity filtration.
- Chlorination.

These combined processes were arranged in combinations and automated sufficiently to permit their 24-hour per day operation while being attended only 8 hours per day (Figure 10). The processes were operated for a full year with the product water being monitored and fed to a small simulated lake to demonstrate the continuous production of high quality water, to compare the alternative chemical treatment and filtration processes for reliability at all times of the year, and to evaluate the behavior of the product water in the simulated lake.

The horizontal flow sedimentation unit was designed for 20 to 50 g.p.m. at overflow rates between 300 and 700 g.p.d. per square foot. The effluent was removed by a submerged launder to avoid floc that might float to the surface. This feature was found to be unnecessary since there was almost no floating floc in a year of operation. A mechanical scraper removed the settled sludge.

Although there was almost no difference in quality characteristics during a year of comparative operation, the sedimentation and dual media gravity filtration systems were selected over the flotation and pressure sand filter systems on the subjective basis of simplicity of operation and general reliability. Alum dosages were increased to 300 mg./l. during the evaluation study to reduce phosphate to its lowest practical minimum. As later tests established, algae growth is not noticeably inhibited at phosphate concentrations above 0.5 mg./l.

Residual turbidity as a function of aluminum sulfate dosage for pilot plant air flotation and horizontal sedimentation tests was averaged over the Phase I and Phase II study (Figure 11). Also, residual phosphate was plotted as a function of alum dosage for pilot plant tests conducted in the Phase I and Phase II study, and averaged for the high alkalinity-high phosphate and low alkalinity-low phosphate periods (Figure 12). These studies show the effectiveness of increasing the alum dosage to 300 mg./l. or more to meet both the turbidity and total phosphate quality criteria for the recreational lake water.

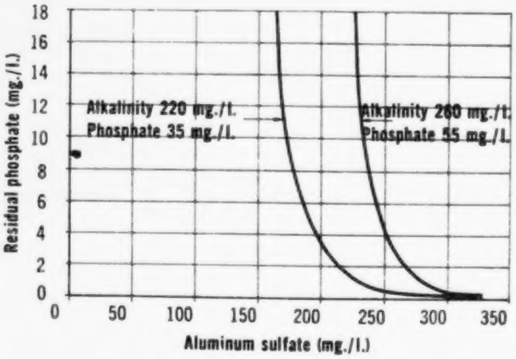
Final selection of processes

The process chosen for converting the oxidation pond water for use in recreational lakes features:

- Coagulation with approximately 300 mg./l. of commercial grade aluminum sulfate and 20 minutes of floc-

Figure 12

Residual phosphate vs. aluminum sulfate prior to filtration (pilot plant)



culation with a typical water treatment plant paddle flocculator.

- Sedimentation with an overflow rate of about 400 g.p.d. per square foot in a horizontal flow tank equipped with sludge scrapers. The sedimentation tank will be covered to prevent wind-generated currents and turbulence.
- Filtration through a dual media gravity filter comprised of 18 inches of 0.5-mm. anthracite over 8 inches of Number 20 sand.
- Chlorination at about 6 to 15 mg./l. Chlorine may be added prior to the sedimentation tank (to reduce crustacean populations reaching the filter) or after the filtration step.
- Minimum 8-hour retention in a chlorine contact chamber.
- Skimmings and sludge from the sedimentation tank and backwash water from the filter will be returned to the inlet of the plant. Most solids will be removed with the primary sludge and digested.

The characteristics of the processed water are compared with those of Pond 3 in Table 1.

A 500,000-g.p.d. facility is being designed to supply the 19-acre recreational lakes. This added supply will prevent dissolved solids in the lake from increasing beyond 700 mg./l. due to evaporation. The water will also be used to irrigate the surrounding park. The capital cost of the plant is estimated at about $150,000. Operating costs are estimated at $184 per million gallons, chemical costs accounting for $70 per million gallons. When expanded to 3-million gallons per day (m.g.d.) capacity, the total cost of capital, operation, and maintenance is estimated at about $150 per million gallons compared to about $200 per million gallons for imported water.

Holding basin studies

Holding basins were constructed to simulate the recreational lake environment. The objectives were evaluation of the process water quality criteria, determination of the chemical and biological behavior of the water, determination of the tolerance of fish for the water, and testing of controlling auxiliary methods of algal regrowth. Selected laboratory tests were also conducted on processed water to determine the relationship between phosphate concentration and algae regrowth.

Three different basins were used: several 1200-gallon, 7.5-foot diameter, 4-foot deep rubber tanks; two 10,000 gallon, 4-foot maximum depth concrete-lined ponds; and an unlined 250,000-gallon, 10-foot deep simulated lake.

During the initial stages of the pilot program, several batch test evaluations of the treated water were conducted in 1200-gallon rubber tanks. Alum and lime were used in several concentrations such that the product water contained between 1 and 5 mg./l. phosphate. Extensive regrowth occurred in each case within two months. Concentrations of *Chlorella* and *Ankistrodesmus* reached 600,000 cells/ml.

In another experiment, phosphate was reduced to 0.2 mg./l. with 300 mg./l. of alum, and the water divided between two 1200-gallon tanks. Dipotassium phosphate was added to one tank to raise the phosphate concentration to 56 mg./l. Regrowth of algae was greatly retarded in the low phosphate water although, after six weeks, both test basins had similar cell counts, turbidity, and appearance.

The high phosphate water produced a great deal more algae in the same time period and accumulated a significant sludge deposit of dead algae on the tank bottom. This test indicated that phosphate reduction would not necessarily influence the actual concentrations of algae to be expected in the recreational lakes but that rates of algae production would be significantly inhibited. As a consequence there should be fewer maintenance problems in the lake due to sludge accumulations and the lake environment should be more suitable for fish.

Draw and fill tests

A long-term experiment consisted of adding alum-treated water with low turbidity and phosphate to the 10,000-gallon pond to maintain about a 40-day detention period. All of the water was filtered and some of it chlorinated. Alum doses varied between 200 and 300 mg./l. (Eight months' data are averaged in Table 2 for the period between November 1964 and June 1965.) Generally pH values, which were between 6.5 and 7.5 for the treated water entering the basin, rose to pH 9 and phosphate was within the 2 mg./l. maximum phosphate quality criterion initially set. The pH rise probably resulted from algae fixing CO_2 and HCO_3 from the water. The water also contained COD and algae concentrations similar to or greater than Pond 3. The dissolved oxygen concentration never reached the intolerably low values observed in Pond 3.

Several other proposed methods of controlling algae regrowth were tested. Lampblack and powdered activated carbon were used without success. In theory, the carbon would adsorb sunlight and prevent light penetration to the algae. However, the carbon failed to stay suspended in the water for more than a few days and no growth inhibition was noticed.

Crustaceans were placed in the water and on two occasions, when the pH was lowered with acid to pH 7, blooms of rotifiers and crustaceans consumed nearly all the algae. However, the crustaceans died and algae growth eventually resumed. Chlorination up to 10 mg./l. is not effective in reducing algae concentrations for more than a few days. Organic algicides and copper compounds controlled certain algae species but they were not effective for all types of algae. Furthermore, they are not effective for any significant time when used at concentrations that would not be harmful to fish.

All of these studies were conducted during the pilot process evaluations when it was still uncertain whether phosphate control could be used to prevent objectional algae growths in the recreational lakes.

Pilot plant operation

In July 1965, the pilot plant began operating continuously and at higher alum dosages. (Chemical data for the 10,000-gallon concrete-lined basin are summarized in Table 2 for the period between September 1965 and April 1966.) Use of 300 mg./l. of alum reduced phosphate residuals to one fifth and algae counts to between one third and one fourth. The predominant algae genera at times changed from the typical green algae usually associated with polluted water (*Scenedesmus, Closteridium*, and the like) to *Nannochloris coccoides Naumann*, very small green algae that have rarely, if ever, been observed in polluted water.

The effect of low phosphate residuals was even more striking in the 250,000 gallon unlined basin (chemical and algae data are shown in Table 2). *Nannochloris coccoides Naumann* was the most prominent and often the only algae genus observed. Because of the extremely small size of *Nannochloris coccoides*, algae counts in the basin did not equal the weight of organic matter in oxidation Pond 3, which averaged about 200,000 counts/ml. The dissolved oxygen levels in this and the smaller 10,000-gallon basin were at least 5 mg./l. or more throughout a year of operation, and the diurnal dissolved oxygen changes were minor compared to those in Pond 3.

Algae will strip most of the remaining orthophosphate from tertiary effluent. Total phosphate concentrations entering the basins averaged 0.25 to 0.3 mg./l., but concentrations in the lakes stayed below 0.05 mg./l. Algae used the phosphate, then settled to the bottom where the phosphate was unavailable. However, extensive decomposition of the sludge layer might cause a sharp increase in phosphate concentration. Low phosphate concentrations (<0.5 mg./l.) limited growth in both holding basins during the continuous pilot plant preferred process studies in which approximately 300 mg./l. alum was used.

Laboratory studies were also conducted to determine if the reduction in phosphate in the pilot plant was in fact inhibiting algae growth. Samples of product water were injected with varying concentrations of phosphate and of other micronutrients (usually Hutner's micronutrient solutions[a]) known to be essential for algae growth. The samples were exposed to continuous fluorescent lighting and stirred with a wrist-action shaker. The tests showed that:

- Some algae growth occurred at phosphate concentrations as low as 0.05 mg./l.
- Phosphate concentrations of 0.1 to 0.3 mg./l. significantly inhibited the rate and density of algae growth.
- Phosphate concentrations below 0.5 mg./l. inhibited algae growth.
- Phosphate concentrations between 1 and 30 mg./l. supported about the same amount of algae growth.

To inhibit algae growth entirely would probably require removal of phosphate to concentrations well below 0.05 mg./l. in the product water. Such removal is economically unfeasible and appears unnecessary because of the apparent suitability of the treated water for the recreational lakes.

TABLE 1

Comparison of Pond 3 and processed water characteristics

Characteristics	Pond 3 water	Product water
pH	8.3	6.7
Turbidity (JTU)	90	4
Total alkalinity (mg./l. as $CaCO_3$)	260	95
Hardness (mg./l. as $CaCO_3$)	80	100
Suspended solids (mg./l.)	75	6
Dissolved solids (mg./l.)	575	575
Chemical oxygen demand (mg./l.)	250	50
Biological oxygen demand (mg./l.)	38	<10
Dissolved oxygen (mg./l.)	0.1–40	7–15
Ammonia–N (mg./l.)	0.1–20	0.1–20
Organic–N (mg./l.)	7–20	1–3
Nitrate–N (mg./l.)	1–4	1–4
Nitrite–N (mg./l.)	0.1–12	0.1–12
Total nitrogen (mg./l.)	7–20	3–20
ABS (mg./l.)	3	3
Phosphate (mg./l.)	40	0.25
Algae (counts/ml.)	200,000	7,000
Confirmed coliform (MPN/100 ml.)	7,900	<1.8
Chlorine residual (mg./l.)	—	0.2–0.5
Sulfate (mg./l.)	60	240

TABLE 2

Comparison of water characteristics in holding basins

Observation period	Water temperature (° C.)	pH	Phosphate (mg./l.)	Algae counts (thousands per ml.)	Predominant algae genera
10,000-gallon holding basin					
Pilot plant process evaluation Nov. 1964–June 1965	11	9.0	0.73	926	Closteridium and Scenedesmus
Continuous pilot plant evaluation Sept. 1965–Apr. 1966	13	8.9	0.12	247	Oscillatoria, Scenedesmus, Nitzschia, and Nannochloris coccoides
250,000-gallon holding basin Sept. 1965–Apr. 1966	15	8.8	<0.05	~2400	Nannochloris coccoides

Fish survival tests

Approximately 125 game fish (25 bass, 50 bluegill, and 50 catfish) were placed in the 10,000-gallon and 250,000-gallon basins for observation. Some 700 Gambusia or mosquito fish and 20 fathead minnows were also added to provide food for the larger game fish. About 90% of the fish are still active in the ponds. A few fish were lost on four different occasions because of specific, but unrelated, causes.

Several fish died when the chlorine residual increased from 0.2 to 1.5 mg./l. in the 250,000-gallon basin. Rising ammonia concentrations in December had eliminated free chlorine residuals and the dosage was increased to 12 to 15 mg./l. to maintain bactericidal control with the less effective

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[a] Hutner's micronutrient solution: The complete solution added 0.76 mg./l. zinc, 0.2 mg./l. boron, 0.14 mg./l. manganese, 0.03 mg./l. iron, 0.04 mg./l. copper, 0.06 mg./l. molybdenum, and 1.75 mg./l. EDTA (ethylenediaminetetraacetic acid). Other solutions with one or more nutrients left out were also used.

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Gerald Stern *is Chemical Engineer and Project Officer, Process Analysis Section, Southern California Field Station, Federal Water Pollution Control Administration, U.S. Department of the Interior, a position he has held since 1962. Previously (1961), he worked for American Potash and Chemical Corp. and (1960–61) Knolls Atomic Power Laboratory–General Electric Corp. Mr. Stern has also been with Goodyear Atomic Corp. (1956–59). He received his B.S. (1955) in chemical engineering from New York University, and has taken graduate courses in chemical engineering, nuclear engineering, and nuclear chemistry at Ohio State University, Rensselaer Polytechnic Institute, and the University of Cincinnati. Mr. Stern is a member of ACS, the Water Pollution Control Federation, and the California Water Pollution Control Association.*

chloramines. The chloramines were so stable, however, that they were not utilized by other organics in the large holding basin. Reduction of the chlorine dose resulted in recovery of the remaining fish.

Again during the colder season, another effect of high ammonia concentrations was observed when the ammonia reached 8 mg./l. and the fish began to exhibit signs of distress. The high ammonia concentration coincided with a rise in pH apparently related to increased algal activity as water temperatures were also climbing.

At higher pH values, the percentage of ammonia in the form of ammonium hydroxide increases substantially and apparently becomes a toxic agent to fish. This problem can be circumvented in the proposed 0.5-m.g.d. tertiary treatment plant by selective storage and withdrawal of water low in ammonia. The ammonia entering the oxidation ponds is readily converted to other nitrogen compounds during periods of warm weather. With an average detention time of 45 to 60 days between the inlet of the oxidation ponds and the point of withdrawal to the tertiary process, ammonia levels do not rise substantially in Pond 3 until some time around December. By April, rising water temperatures cause a sudden drop in ammonia concentrations throughout the oxidation ponds. By filling oxidation Pond 3 to capacity prior to the rise in ammonia concentrations and then bypassing around it during the winter, there will be low ammonia content water for the recreational lakes throughout the year.

During the latter part of the cold season in 1966, the dissolved oxygen level fell below 3.5 mg./l. throughout the 250,000-gallon simulated lake. The water had a brownish appearance and algae counts decreased to 50,000 cells/ml. Some fish were in distress. Artificial aeration was provided and the dissolved oxygen level increased to saturation within three days. The brownish color was probably caused by a bacterial bloom that used up more oxygen than was being produced by the few algae present.

This experience emphasized that any biological system is subject to unpredictable changes and that blooms of bacteria, crustaceans, or algae may result in occasional oxygen depletion in any recreational lake using reclaimed water. Some method for supplementing the oxygen content of recreation lake water should, therefore, be provided for emergency use.

Conclusions

Turbidity, algae, and phosphate reduction of conventional oxidation pond water can be economically achieved. A practical tertiary process has been developed to supply boating and fishing lakes near Lancaster, Calif. The process selected and piloted at 30 g.p.m. features treatment with 200 to 300 mg./l. of alum, sedimentation, filtration, and chlorination. Phosphate is consistently reduced from 40 mg./l. to less than 0.5 mg./l. and algae counts from several hundred thousand to a few thousand per milliliter. Product water has been evaluated in the laboratory and in large simulated lakes stocked with fish. Algae slowly grow to high concentrations in the water but the rate of biological activity does not deplete the oxygen supply, or interfere with fish life, or cause nuisances. Laboratory stimulated growth tests show that algae growth rates are inhibited at phosphate concentrations less than 0.5 mg./l. and almost stop at less than 0.05 mg./l. The ecology of the simulated lake is typical of a fresh water environment supplied with limited nutrients.

ADDITIONAL READING

Cohen, J. M., "Improved Jar Test Procedure," *J. Am. Water Works Assoc.* **49** (11), pp. 1425–31 (1957).

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ADVANCED WASTE TREATMENT AND
WATER REUSE SYMPOSIUM

Session #2, Tuesday, February 23

Moderators: Minnesota and Michigan Water Pollution Control Agency Representatives

12:45 P.M. Oxygen Aeration for Secondary Treatment

Richard C. Brenner, Research Sanitary Engineer
Biological Treatment Research
Advanced Waste Treatment Research
Laboratory, EPA, Cincinnati

1:30 P.M. Biological Nitrification and Denitrification

Edwin F. Barth
EPA, Cincinnati

2:15 P.M. The Problem of Viruses in Waste, Renovated and Other Waters

Dr. Gerald Berg, Chief
Virology Research
Advanced Waste Treatment Research
Laboratory, EPA, Cincinnati

3:00 P.M. to 3:15 P.M. Coffee Break

3:15 P.M. Phosphorous Removal in Conventional Treatment

Edwin F. Barth
EPA, Cincinnati

4:00 P.M. Combined Biological-Chemical Treatment for Phosphorous Control

Richard C. Brenner
EPA, Cincinnati

CONTENT

"Oxygen Aeration for Secondary Treatment"

Richard C. Brenner
Research Sanitary Engineer
EPA, Cincinnati

"Nitrogen Removal from Wastewater"

Advanced Waste Treatment Research Laboratory
EPA, Cincinnati

"Phosphorous Removal in Conventional Treatment"

Edwin F. Barth, Chemist
EPA, Cincinnati

"Combined Biological-Chemical Treatment for Control of Phosphorous"

Richard C. Brenner
Research Sanitary Engineer
EPA, Cincinnati

Additional Papers Included for Future Reference

"Bacterial and Viral Control and Water Reuse (for Domestic Purposes)"

Harold W. Wolf, Dr. P. H.

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ADVANCED WASTE TREATMENT AND
WATER REUSE SYMPOSIUM

HIGH PERCENTAGE OXYGEN AERATION FOR SECONDARY TREATMENT

Richard C. Brenner
Advanced Waste Treatment Research Laboratory
U.S. Department of the Interior
Federal Water Quality Administration
Ohio Basin Region
Cincinnati, Ohio

HIGH PERCENTAGE OXYGEN AERATION FOR SECONDARY TREATMENT

Richard C. Brenner

INTRODUCTION

The concept of using oxygen enriched air or pure oxygen as the aeration gas in the activated sludge process has intrigued investigators for over twenty years. In 1948, Pirnie (1) suggested a method of predissolving oxygen in wastewater prior to activated sludge treatment. The following year, Okun expanded on Pirnie's model and developed his "bio-precipitation system" which utilized an integral upflow bio-reactor and clarifier (2). Later, Budd and Lambeth (3) evaluated Okun's "bio-precipitation system" in a large pilot plant and reported an oxygen absorption efficiency of 20-25% with power requirements equal to the conventional diffused air technique.

The inherent characteristic of oxygen gas which has created so much interest over the years is its high partial pressure. At atmospheric pressure, the partial pressure of pure oxygen gas is 760 mm of mercury while the partial pressure of oxygen in air is only 160 mm of mercury. Thus, the driving force for dissolving oxygen in water is 4.75 times higher with pure oxygen.

In a "state-of-the-art" review, McKinney and Pfeffer (4) concluded that if the potential of high percentage oxygen aeration could be harnessed in an economically realistic fashion, the following benefits to the activated sludge process could conceivably be achieved:

1. Reduction of power required per unit of oxygen transferred.
2. Reduction in, or elimination of, periods of zero dissolved oxygen concentration.
3. Increased rate of stabilization of organic material.
4. Ability to operate high rate systems with substantial increases in organic loading where oxygen is not limiting.
5. Increased capacity of organically overloaded plants without need for additional aerator capacity.
6. Reduction in plant size and capital investment.

Until recently, however, serious consideration of oxygen aeration has not progressed past the laboratory and pilot plant stages. The basic problem has been the lack of an effective oxygen gas contacting system with a built-in high overall utilization factor. Since oxygen is an expensive commodity and since conventional aeration techniques would vent the major fraction of the input oxygen to the atmosphere, the widespread impression has been that oxygen aeration must, of necessity, be significantly more costly than air aeration. This rationale led Okun and Lynn (5) to conclude in 1956 that the substitution of oxygen for air or the use of oxygen enriched air in standard activated sludge aeration tanks would not be economically feasible because of the poor oxygen transfer efficiency inherent in diffuser and tank design.

Okun and Lynn's position as stated above has proven to date to be partially correct. Certainly, the requirements for an economical oxygen contacting system are such that conventional air aeration technology is entirely inadequate. However, what Okun and Lynn did not forsee was the covering of standard activated sludge tanks to form a gas tight structure. An efficient oxygen aeration system (which will be discussed at length in the following sections) has been developed which utilizes covered longitudinal aeration tanks. The development of an efficient open tank oxygen aeration system also appears to be within the realm of feasibility.

STIMULATION OF RENEWED INTEREST IN OXYGEN AERATION

In early 1968, the Linde Division of Union Carbide approached the FWQA concerning a new oxygen gas contacting concept developed in their laboratories. According to Union Carbide, this concept satisfied all the basic requirements of a cost effective oxygen dissolution system:

1. High percentage oxygen absorption and utilization (i.e. >90%).
2. High overall energy transfer efficiency as compared with air aeration systems.
3. Efficient liquid mixing sufficient to maintain uniform mixed liquor composition.
4. Simple design with low maintenance costs.
5. Low capital investment.

Continuing discussions culminated in the awarding of a $528,000 FWQA Research and Development Contract to Union Carbide in October, 1968. The contract was carried out at the Batavia, New York Treatment Plant. Under the terms of the contract, the performance and economics of the

Linde oxygen aeration (oxygenation) system were evaluated and compared with those of a parallel air aeration system from May to November of 1969.

CONFIGURATION AND OPERATION OF OXYGENATION SYSTEM

A schematic respresentation of the Union Carbide oxygenation system is shown in Figure 1. A gas tight cover on top of the aeration tank is utilized to prevent oxygen leakage to the atmosphere. Baffles are installed to form a series of staged comartments down the length of the aerator. Each compartment is equipped with a submerged turbine-rotating sparger combination. A recirculating gas compressor is located on top of or adjacent to each compartment.

Liquid waste and oxygen gas flow are cocurrent with both being introduced to the head of the aeration tank. Recycle sludge also enters at this point. Oxygen gas fed to the first compartment is compressed and forced down the hollow shaft of the submerged turbine and out through the rotating sparger. The fine bubbles thus created are dissolved in the mixed liquor. The propeller blades furnish bulk fluid mixing and create a downward pumping action which provides longer residence time for the bubbles.

Oxygen which is not utilized by microorganisms vents back into the head space of the first compartment. Here it is either compressed and recirculated again or, due to the slight positive pressure of 1-3" water column maintained in the gas phase, moves through an appropriately sized opening in the baffle wall to be used in the second compartment by the same technique. In like fashion, vent gas from each compartment is used in the succeeding compartment until the end of the aeration tank is reached. A small exhaust stream (10-20% of the input gas stream) vents to the atmosphere from the last compartment. It should be pointed out that the gas stream is picking up reaction products and inert gases as it proceeds down the compartments. The exhaust gas is, therefore, about 50% oxygen, and 50% carbon dioxide, nitrogen, argon, etc. The small exhaust rate and the reduced composition of oxygen in the exhaust gas combine to yield an overall oxygen utilization of 90-95%.

SCOPE OF THE BATAVIA PROJECT

The nominal capacity of the Batavia plant is 2.5 mgd. As indicated in Figure 2, the plant consists of two parallel and identical 1.25 mgd aeration tanks each with its own final clarifier and separate return sludge line. Each aeration tank has two passes. There are no primary settlers at Batavia. Raw wastewater is screened, degritted, and comminuted before entering the aerators. Waste activated sludge is handled in sequence by gravity thickening, two stage anaerobic digestion, vacuum filtration, and trucking to disposal. Treated effluent is chlorinated before discharge to Tonawanda Creek.

The oxygenation system was installed in one of the two aeration tanks. A temporary steel deck cover was bolted and sealed to the top of the oxygen aerator for convenience and economy. Baffle installations were also designed for short term use. Six compartments or stages were utilized, three in each pass. Oxygen was stored on site in the liquid form and then vaporized before introduction to the oxygen aerator. Oxygen feed rate to the first compartment was automatically controlled by maintaining a constant vapor pressure in the gas space of 1-3" of water. Dissolved oxygen level was also automatically controlled, but in the first compartment only where oxygen demand was highest. This was accomplished with a dissolved oxygen probe and a feedback signal which controlled the rate of recirculation of the first stage compressor.

The existing diffused air system was left intact in the second aerator for the comparative studies. Aeration capacity of this train was increased with 50% additional laterals and diffusers. Compressed air was supplied to the air train by the plant's rotary blowers.

The experimental program was segmented into three phases; a conservative phase, a medium stressed phase, and a highly stressed phase. To take advantage of dry weather flow, the highly stresed phase (Phase II) was conducted in the summer immediately after the conservative phase (Phase I). The medium stressed phase (Phase III) was conducted last. The modes of operation of the three phases are given in Figures 3, 4, and 5.

In Phase I, the average plant flow of 3.88 mgd was divided evenly as nearly as possible between the two trains. All six compartments of the oxygen train and both passes of the air train were utilized. The intended aeration detention time in each train based on raw flow was 6 hours, but the high spring flows reduced this to about 4 hours. Seven weeks of comparative data were collected in this phase.

All of the plant flow was directed to the oxygen train in Phase II. To further stress the system, three of the six stages were taken out of service. Both clarifiers and air lift pumps were utilized to separate and return the sludge. The intended and actual aeration time averaged 1.5 hours. Seven weeks of data were also collected in this phase.

The plant flow was again split between the two trains in Phase III. Three of the six oxygen compartments and one pass of the air train were taken out of service to achieve a medium loaded condition. The intended aeration detention time of 3 hours was realized with the air train. The detention time of the oxygen train averaged somewhat lower at 2.8 hours. Nine weeks of comparative data were collected in this phase.

SUMMARY AND DISCUSSION OF BATAVIA PERFORMANCE DATA

Comparative performance data generated on the Batavia project are summarized in Figures 6 through 9. Each of the data figures will be discussed in order below:

Figure 6: Excellent operation and surveillance were given to both systems throughout. The degree of operator attention was 2-3 times that normally expended at treatment plants. Well-trained Union Carbide technicians operated the experimental facilities during the week; Batavia plant personnel were utilized on the weekends.

It was decided to operate the air train in Phase I at a mixed liquor suspended solids (MLSS) concentration typical of other conventional air systems of approximately 2500 mg/l. MLSS in the oxygen train in Phase I were kept at 40-50% of the maximum concentration which it was anticipated could be carried in later phases, or about 3000 mg/l. In order to treat effectively under the high rate loading conditions of Phase II, oxygen train MLSS were raised to the maximum sustainable level of 7000 mg/l. This upper limit was not due to oxygen transfer limitations but to restricted clarification capacity. In Phase III, oxygen train MLSS were kept high at about 6200 mg/l while air train MLSS were raised to the maximum point (3700 mg/l) at which some measurable dissolved oxygen (DO) could still be maintained in the mixed liquor. Above 3700 mg/l, DO fell to zero and septic odors commenced.

Sludge settled very well in both systems. Sludge volume index (SVI) varied from 35-65 for the oxygen system and 65-75 for the air system. In both systems, SVI decreased with increasing MLSS. In all probability, the extremely good settling characteristics were due in part to the absence of primary sedimentation.

Recycle sludge concentration was also higher than in most plants. For the air system, the concentration ranged from 1.5-1.6%. The range for the oxygen system was 1.8-3.0%. Recycle sludge concentrations in the oxygen system as high as 4% were noted in Phase II during the day. The installation of a larger flow meter in the oxygen train sludge recycle line to handle the return from both clarifiers in Phase II adversely affected oxygen slydge recycle concentration in Phase III. Return flow had to be kept artificially high in Phase III to maintain sufficient head loss to operate the oxygen air lift pump, thereby thinning out the sludge. It is felt that with more conservative overflow rates and variable speed return pumps, oxygenated sludge can be thickened routinely to 3-4% in the final clarifier. At these concentrations, it is doubtful if the waste oxygenated sludge would have to undergo separate thickening before digestion or direct vacuum filtration.

Mixed liquor DO averaged 8-9 mg/l in the oxygen train in all phases. Variations in DO profile down the stages were insignificant. The low oxygen effluent DO in Phase II indicates that some treatment was still going on in the final clarifier due to the high organic loading. Air train mixed liquor DO was low in both Phases I and III. No measurable DO was present at the head of the air aerator in either phase, despite abnormally high system air input rates of 3-4 ft^3/gal of sewage treated. The poor oxygen transfer efficiency of disc diffusers is readily apparent.

Figure 7: As anticipated, BOD and suspended solids (SS) removals were good for both systems in the conservative Phase I. Effluent COD values of 73 mg/l for the oxygen train and 84 mg/l for the air train are considered acceptable for aerators treating unsettled sewage. Volumetric organic loadings were in the upper portion of the range for which most conventional plants are designed. Overall performance in Phase I was slightly better for the oxygen train.

One of the most encouraging aspects of the entire project was the acceptable level of performance of the oxygen system under the high rate conditions of Phase II. Effluent BOD, COD, and SS values were not as low as in Phase I but still within the range of good secondary treatment. The volumetric organic loading was 3-1/2 times higher than in Phase I. The ability of the oxygen system to operate satisfactorily under much higher organic loading rates than normally associated with air systems was clearly established in this phase.

Wastewater feed strength was at its highest in Phase III. Some problems were encountered with the flow splitting device to the two aerators with the result that about 10% more flow was directed to the oxygen train. Performance of the oxygen train was excellent. Percent removals of BOD, COD, and SS were the highest of any of the three phases although absolute effluent concentrations were slightly higher than in Phase I. The air system began to show signs of performance deterioriation in Phase III. There was zero DO in the air train effluent. Effluent BOD and SS concentrations were about double and effluent COD 30% higher than those of the oxygen system. Volumetric organic loadings were more than twice as high as in Phase I.

Somewhat unexpected was the magnitude of the reduced quantities of waste activated sludge generated throughout the study in the oxygen system. Disregarding the data point for the oxygenation system in Phase III (which is extremely out of balance with the other two phases), 30-40% less sludge was wasted from the oxygen system than from the air system. Considering that sludge handling and disposal account for 1/2 to 2/3 of total annual treatment costs, this factor assumes significant economic importance. From the above data, it is evident that oxygenation is a highly oxidative process.

Figure 8: The two principal conclusions which can be drawn from these data are that phosphorus removal was in the range normally associated with biological uptake for both systems and that neither system nitrified. For the two comparative phases, phosphorus removal averaged 5% higher for the air system. This is to be expected as the oxygen system "burned up" more sludge and thereby, in all likelihood, solubilized greater quantities of stored cellular phosphorus. Only the very beginnings of incipient nitrification were noticed in either unit. This was the one disappointing feature of the Batavia oxygen train performance, notwithstanding that prevailing temperatures and organic loading rates were not exactly conducive to nitrification.

Figure 9: High percentage utilization of feed oxygen was achieved at greater than 90% in each phase. The average composition of oxygen in the exhaust gas for the project was 51%.

Power consumption for oxygen dissolution is broken into two categories, liquid mixing and gas recirculation. Power consumption for liquid mixing is roughly proportional to the gallons of mixed liquor under oxygenation while gas recirculation power is highly dependent on waste strength and the total pounds of BOD or COD entering the system. It should be pointed out that the oxygen compressors used on this project were substantially oversized for experimental flexibility and that the power requirements shown here for gas recirculation represent only that fraction of gas which was actually put through the turbine shaft. Power consumed in bypassing unneeded gas is not included. This is a legitimate technique as in a non-experimental situation, compressors are sized more closely to design conditions and variable speed compressors are used in the high oxygen demand stages.

Abnormally high aeration rates were required in the air train in an attempt to cope with the high organic strength of the unsettled wastes. This resulted in an air blower power requirement of 5-6 times that for oxygen dissolution. For a true comparison of total power requirements, however, the power consumed in on-site oxygen gas production should be added to that used in oxygen dissolution. Since liquid oxygen was used in this project, no such totals are possible here. In general, though, power requirements for oxygen generation will range from 1.3 to 1.7 times greater than the power required for oxygen dissolution. Even so, a considerable reduction in overall power costs with oxygen is indicated..

ECONOMIC PROJECTIONS

The following assumptions are used in conjunction with the Batavia data in arriving at comparative cost curves for the two systems:

1. Raw sewage pumping and metering, preliminary treatment (including screening, comminution, and grit removal), primary sedimentation, service and administration buildings, and outside piping and appurtenances are costed in equally for both systems.

2. Raw wastewater BOD is assumed to be 200 mg/l with 35% removed in the primary settlers.

3. Equal amounts of primary sludge are produced in each system at the rate of 860 lb/mil gal treated. Waste activated sludge quantities for the two systems are in accordance with Figure 10 which was derived from the Batavia data. Phase III oxygen system waste sludge data were not used in constructing Figure 10.

4. MLSS concentration for the oxygen and diffused air systems are assumed to be 6800 mg/l and 2500 mg/l, respectively, with 65% of the sludge being volatile in each case.

5. Based on assumption 4 and using Figure 11 which was compiled from the Batavia work, 1 hour of oxygen aeration will provide equivalent treatment to 3 hours of air aeration.

6. For each size plant costed, four aeration detention points (based on raw flow) are plotted and compared. These points are 1.0, 1.33, 1.66, and 2.0 hours for oxygen and 3.0, 4.0, 5.0, and 6.0 hours for air. Percent BOD removal associated with each set of comparative points is given in Figure 12. Volumetric organic loadings and food to biomass ratios for the various aeration detention times are shown in Figure 13.

7. Due to the variance in oxygen transfer capability of different types of air diffusers, three air aeration rates are costed for the air system including 0.8, 1.6, and 2.4 ft^3/gal wastewater treated.

8. Oxygen supply varies from 1.0 to 1.22 lb/lb BOD removed as oxygenation detention time increases from 1.0 to 2.0 hours. Average oxygen utilization is 91%.

9. Mixed liquor DO is assumed at 8 mg/l for the oxygen system and 1 mg/l for the air system.

10. Sludge recycle oncentrations are assumed to be 40,000 mg/l and 10,000 mg/l, respectively, for oxygen and air systems.

On these bases, average sludge recycle rates in the same order are 20% and 33% of influent flow. Spare recycle pump capacity of 100% average recycle flow is costed in for each system.

11. Oxygen is supplied to the oxygenation system by on-site generation of oxygen gas. For plant sizes up to approximately 30 mgd, the generation plant would be of the air separation type. Above 30 mgd, a cryogenic plant is used. Union Carbide has made a managerial decision that municipalities, if they so choose, may purchase the plant outright rather than purchasing liquid or gaseous oxygen across the fence. The savings to the municipality in $/ton of oxygen by buying the plant instead of the commodity itself are evident in Figure 14. The bottom curve is used in the cost projections given herein.

12. Total oxygenation system and air aeration system power requirements for the four plant sizes (1, 6, 30, and 100 mgd) used later in the comparative cost curves are indicated in Figures 15 and 16, respectively. Power requirements for the oxygen system are a function of oxygenation detention time; those for the air system are a function of air aeration rate. Reserve air system blower capacity of 100-25% is provided as plant size increases from 1 to 100 mgd. Backup capacity for the oxygen generators includes a spare air compression unit for the 1, 6, and 30 mgd plant sizes and a 24-hour on-site supply of liquid oxygen for the 100 mgd plant.

13. Sludge disposal for the 1 mgd plant is assumed to be by anaerobic digestion followed by sludge drying beds, for the 6 mgd plant by anaerobic digestion, vacuum filtration, and cake disposal by land fill, and for the 30 and 100 mgd plants by anaerobic digestion, vacuum filtration, and incineration. Both primary and waste activated sludges were costed into the sludge disposal estimates.

14. A gravity thickener is costed in for the air system to thicken waste activated sludge from 1 to 4% prior to anaerobic digestion. Since it is assumed that waste activated sludge is drawn from the oxygen system at 4%, a thickener is not needed for the oxygen system.

15. Secondary clarifiers are sized at 800 gal/day/ft^2 for both systems at a liquid depth of 10 ft for the air system and 11 ft for the oxygen system.

16. Cost curves and formulae for construction and operation of conventional plant units and for sludge thickening and disposal were supplied to Union Carbide by FWQA. Cost data for oxygen generation, dissolution, and control equipment and for oxygenation tank covers and baffles were furnished by Union Carbide. Amortization was calculated at 5-1/2% over 25 years. Construction costs were determined using January, 1970 indices.

The above assumptions have been compiled into four comaparative cost estimate curves for 1, 6, 30, and 100 mgd plant sizes. These cost estimates are given in Figures 17 through 20 as total treatment costs in ¢/thousand gallons treated. The costs reflect debt retirement and operation and maintenance of all plant facilities including sludge disposal equipment, i.e. the total cost associated with running a combined primary-secondary treatment plant with the exception of effluent chlorination.

As can be seen in Figure 17, there is no cost advantage with oxygen at 1 mgd. At the 6 mgd scale, there is an estimated cost savings with oxygen ranging from 11-15% when compared with an air aeration rate of 0.8 ft^3/gal and 17-21% when compared with an air aeration rate of 1.6 ft^3/gal. For the 30 mgd plant, estimated cost reductions with oxygen range from 16-20% and 21-27% for the comparisons with 0.8 and 1.6 ft^3/gal, respectively. At 100 mgd, the estimated cost reductions with oxygen are 17-25% and 22-29% compared to the same air aeration rates, respectively.

It is emphasized that the cost data shown in these figures are estimates only and that individual plant costs may vary widely with locale. Further, it is realized that air aeration equipment vendors, particularly those that manufacture surface and turbine aerators, may disagree with the air system design used in the cost comparisons. Nevertheless, it is believed the air system design so presented reflects conventional plug flow designs as they are approved by state regulatory agencies and used in the majority of the municipalities in this country. Likewise, the oxygen system design presented herein reflects the manner in which it is contemplated such a system would be designed for a municipal installation.

APPLICATIONS OF THE OXYGEN AERATION PROCESS

It is felt the results of the Batavia project justify consideration of the oxygen aeration process for all new secondary plants and primary plants expanding to secondary treatment. In this type of application, the City of Detroit, Michigan has solicited bids with State of Michigan and FWQA Construction Grants approval for a 300 mgd oxygen aeration module for its new METRO plant. Also, the State of New York has approved plans for a 15 mgd oxygen aeration system at New Rochelle, New York.

The oxygenation process also appears to be equally applicable to upgrading existing overloaded activated sludge plants. It is believed that the aeration capacity of many overloaded systems can be increased 2-3 times without acquiring additional land for more aeration tanks. To test this hypothesis, the FWQA has recently awarded a $1.5 million Demonstration Grant to New York City for conversion of one existing 20 mgd bay of its Newtown Creek Plant to oxygen. The plant is currently accomplishing about 40% BOD removal with 1-1/2 hours of air aeration. It is hoped the removal efficiency of the test bay will be improved to 90% with oxygen. The test bay is scheduled to be on stream by late fall, 1971, and the grant evaluation will then continue for one year.

Other cities where Union Carbide is conducting non-federally sponsored pilot plant studies include Cincinnati, New Orleans, Louisville, Euclid, Ohio, Middlesex, New Jersey, and Speedway, Indiana.

AREAS NEEDING FURTHER INVESTIGATION

Questions remain to be answered concerning the following aspects of aeration with oxygen:

1. Nitrification: As mentioned previously, the Batavia experiments did not demonstrate nitrification. Definite criteria need to be established at which nitrification with oxygen can be guaranteed. Experiments are currently underway at the FWQA Blue Plains pilot plant in which a two-step oxygenation flow regime with intermediate clarification is being utilized to study this problem.

2. Sludge handling and disposal: The high solids concentrations and reduced quantities of waste activated sludge generated with oxygenation suggest several alternative approaches to anaerobic digestion for sludge disposal. Because of the highly oxidized condition of the sludge, several days of aerobic digestion may be sufficient to stabilize the sludge for land spreading. It may also be possible to dewater waste sludge directly from the clarifier underflow without any type of prior digestion. A follow-on contract for $208,714 was awarded to Union Carbide for further studies at Batavia during the summer and fall of 1970, a part of which included direct vacuum filtration of waste oxygenated sludge, aerobic stabilization of waste oxygenated sludge, and vacuum filtration of aerobically stabilized sludge. The final report from this follow-on contract will be available around May, 1971.

3. Carbon dioxide buildup in mixed liquor: Because of the closed aeration system, carbon dioxide (CO_2) tends to build up in solution in an oxygen aerator. At Batavia, this caused a drop

in mixed liquor pH of 0.5-0.7 units in all phases. Thus, it can be speculated that for stronger wastes more CO_2 will be produced and the drop in pH could be large enough to endanger the sludge culture.

Air Reduction Company is currently conducting a pilot study at the Greensboro, North Carolina South Buffalo Plant. In addition to domestic wastes, this plant receives all of the City's textile mill wastes. Organic strength of the combined municipal wastewater is very high. The Air Reduction oxygen aeration system uses a large diameter PVC pipe reactor to replace the standard open or closed aeration tank. Oxygen injection to the pipeline can be single or multi point as needed. The advantage of this system is that it requires no moving parts other than the prime mover for oxygen dissolution. If the pilot studies indicate the high strength wastewater can be effectively treated with oxygen, it is contemplated that the City of Greensboro will apply to FWQA for a Demonstration Grant, during which CO_2 buildup would be further investigated.

4. Effect of primary sedimentation: The effect of primary sedimentation on the oxygen process, particularly on sludge settling characteristics is not known. At this time, a recommendation for or against primary sedimentation in conjunction with oxygen aeration can not be made. The Newtown Creek Plant does not have primary settlers, so the answer to this question will have to come from privately funded studies or future federal grant projects.

CONCLUSIONS

1. Oxygen aeration is a highly oxidative process which operates effectively as a high rate system.

2. Significant reductions in waste activated sludge production are indicated with oxygen aeration.

3. Oxygen sludge settles and thickens well. It may be possible to dewater waste sludge from this process directly from the clarifier underflow.

4. The oxygen aeration process is applicable to both new and existing overloaded activated sludge plants.

5. Cost reductions in total treatment costs of 15-30% appear feasible for new plants with oxygen. Percent savings will be greater for conversion of existing overloaded plants to oxygen.

6. Several aspects of the oxygen aeration process require further investigation to maximize its full potential for waste treatment.

APPENDIX

All of the figures utilized in this discussion are reproduced directly from the final report of the Batavia project (6). Copies of this report can be obtained by writing to either of the following:

Planning and Resources Office
Office of Research and Development
Federal Water Quality Administration
U. S. Department of the Interior
Washington, D. C. 20242

Advanced Waste Treatment Research Laboratory
Federal Water Quality Administration
U. S. Department of the Interior
4676 Columbia Parkway
Cincinnati, Ohio 45226

Attention: Mr. R. C. Brenner

REFERENCES

1. Pirnie, M., Paper presented at 21st Annual Meeting of Federation of Sewage Works Associations, Detroit, Michigan (October 1948).

2. Okun, D. A., "System of Bio-Precipitation of Organic Matter from Sewage," Sewage Works Journal, 21 (1949).

3. Budd, W. E. and Lambeth, G. F., "High Purity Oxygen in Biological Sewage Treatment," Sewage and Industrial Wastes, 29, 253 (1957).

4. McKinney, R. E. and J. T. Pfeffer, "Oxygen Enriched Air for Biological Waste Treatment," Water and Sewage Works (October 1965).

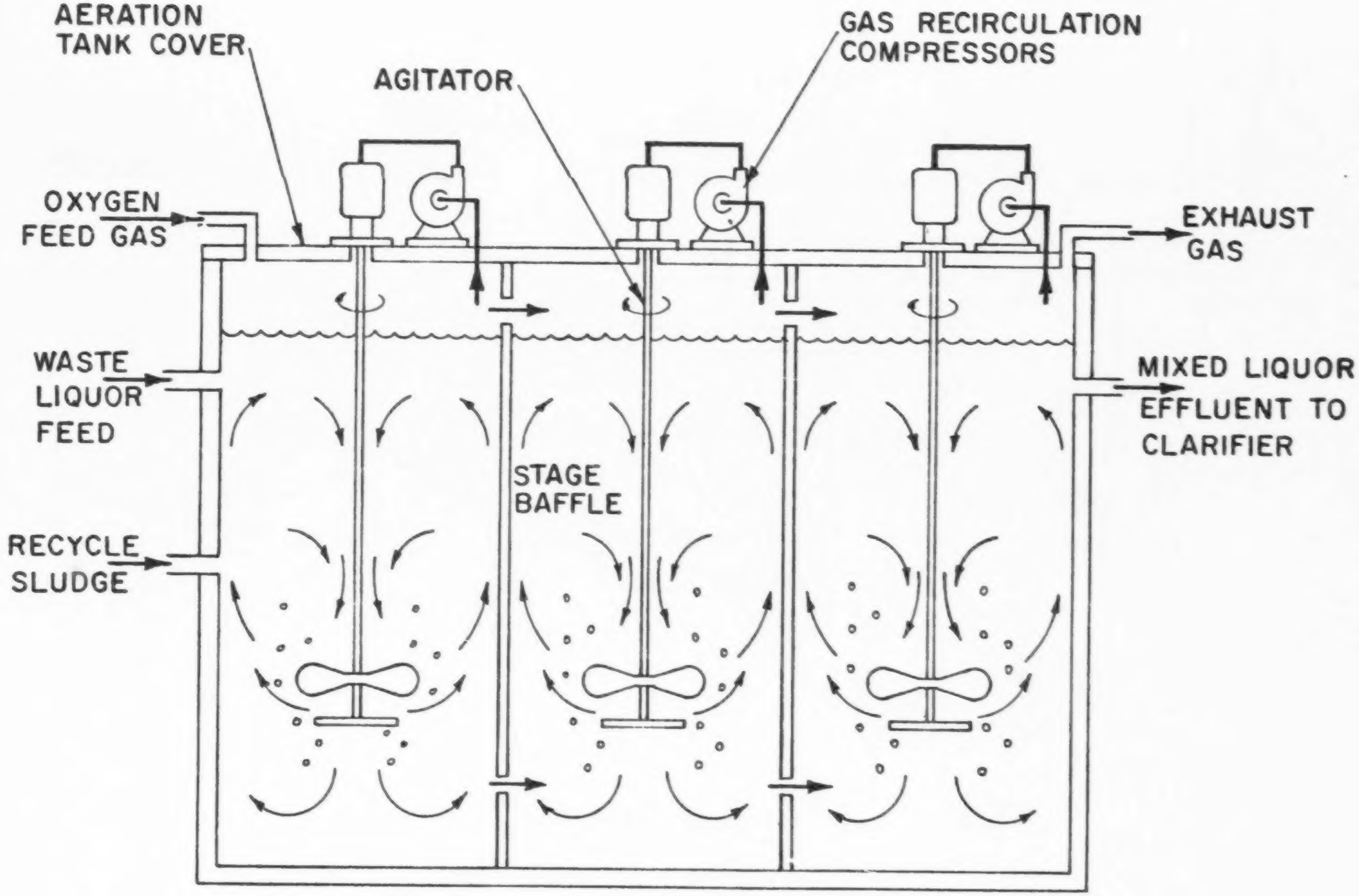
5. Okun, D. A. and Lynn, W. R., "Preliminary Investigation into the Effect of Oxygen Aeration on Biological Sewage Treatment," Biological Treatment of Sewage and Industrial Wastes: Volume I, Aerobic Oxidation, Reinhold Publishing Corp., New York (1956).

6. Albertsson, J. G., McWhirter, J. R., Robinson, E. K., and Vahldieck, N. P., "Investigation of the Use of High Purity Oxygen Aeration in the Conventional Activated Sludge Process," FWQA Research Contract No. 14-12-465, Program No. 17050 DNW (May 1970).

90<

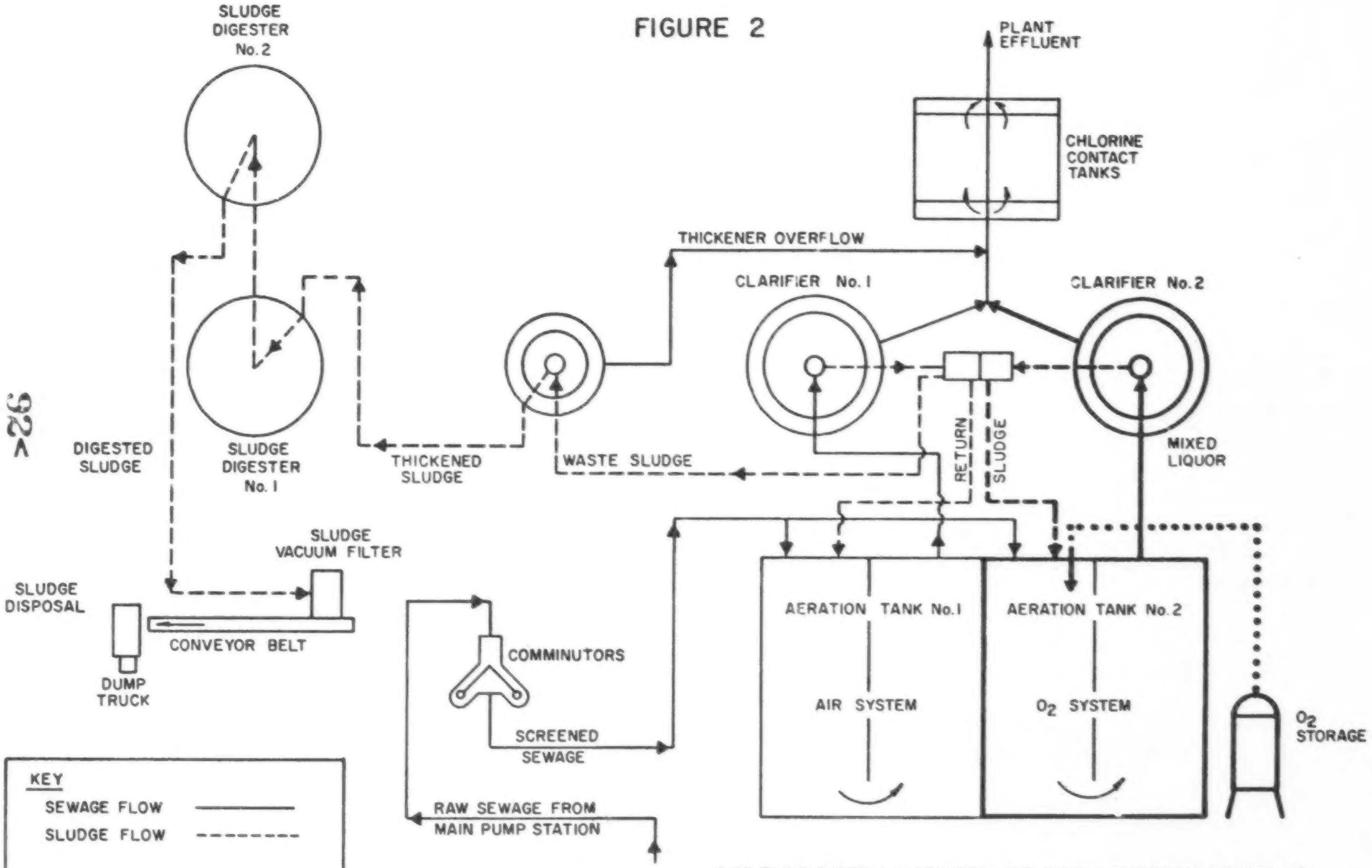
FIGURE I

SCHEMATIC DIAGRAM OF MULTI-STAGE OXYGENATION SYSTEM


AERATION TANK COVER
GAS RECIRCULATION COMPRESSORS
AGITATOR
OXYGEN FEED GAS
EXHAUST GAS
WASTE LIQUOR FEED
MIXED LIQUOR EFFLUENT TO CLARIFIER
STAGE BAFFLE
RECYCLE SLUDGE

WATER POLLUTION CONTROL PLANT, CITY OF BATAVIA, N.Y.
SCHEMATIC FLOW DIAGRAM

FIGURE 2


SLUDGE DIGESTER No. 2
SLUDGE DIGESTER No. 1
DIGESTED SLUDGE
THICKENED SLUDGE
WASTE SLUDGE
THICKENER OVERFLOW
PLANT EFFLUENT
CHLORINE CONTACT TANKS
CLARIFIER No. 1
CLARIFIER No. 2
RETURN
SLUDGE
MIXED LIQUOR
SLUDGE VACUUM FILTER
SLUDGE DISPOSAL
CONVEYOR BELT
DUMP TRUCK
COMMINUTORS
SCREENED SEWAGE
RAW SEWAGE FROM MAIN PUMP STATION
AERATION TANK No. 1
AIR SYSTEM
AERATION TANK No. 2
O_2 SYSTEM
O_2 STORAGE

KEY

SEWAGE FLOW ————

SLUDGE FLOW --------

DESIGN POPULATION 25,000
AVG FLOW: 2.5 MIL. GAL./DAY
MAX. FLOW: 6.25 MIL. GAL./DAY

THIS FLOW SHEET ILLUSTRATES THE WASTE TREATMENT PLANT AT BATAVIA, NEW YORK. BECAUSE THIS IS A DUAL-TRAIN SYSTEM, ACCURATE COMPARISONS BETWEEN AIR AERATION AND THE OXYGENATION SYSTEM ARE POSSIBLE USING THE SAME WASTE WATER. THE PART OF THE PLANT OUTLINED IN HEAVY LINES HAS BEEN ADAPTED FOR OXYGEN USE.

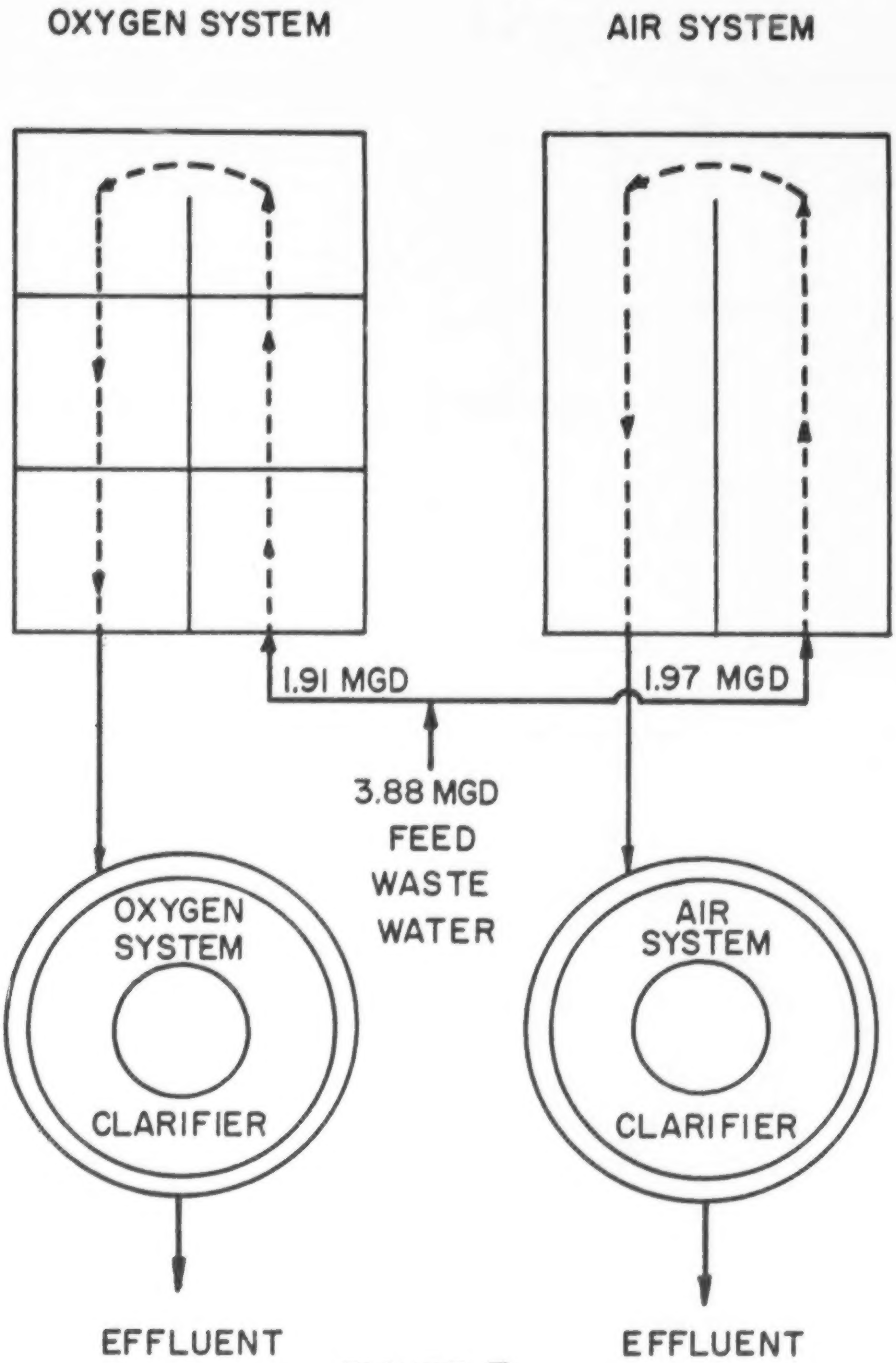

OXYGEN SYSTEM
AIR SYSTEM
1.91 MGD
1.97 MGD
3.88 MGD
FEED
WASTE
WATER
OXYGEN
SYSTEM
CLARIFIER
AIR
SYSTEM
CLARIFIER
EFFLUENT
EFFLUENT

FIGURE 3

BATAVIA TREATMENT SYSTEM

PHASE # 1 OPERATION

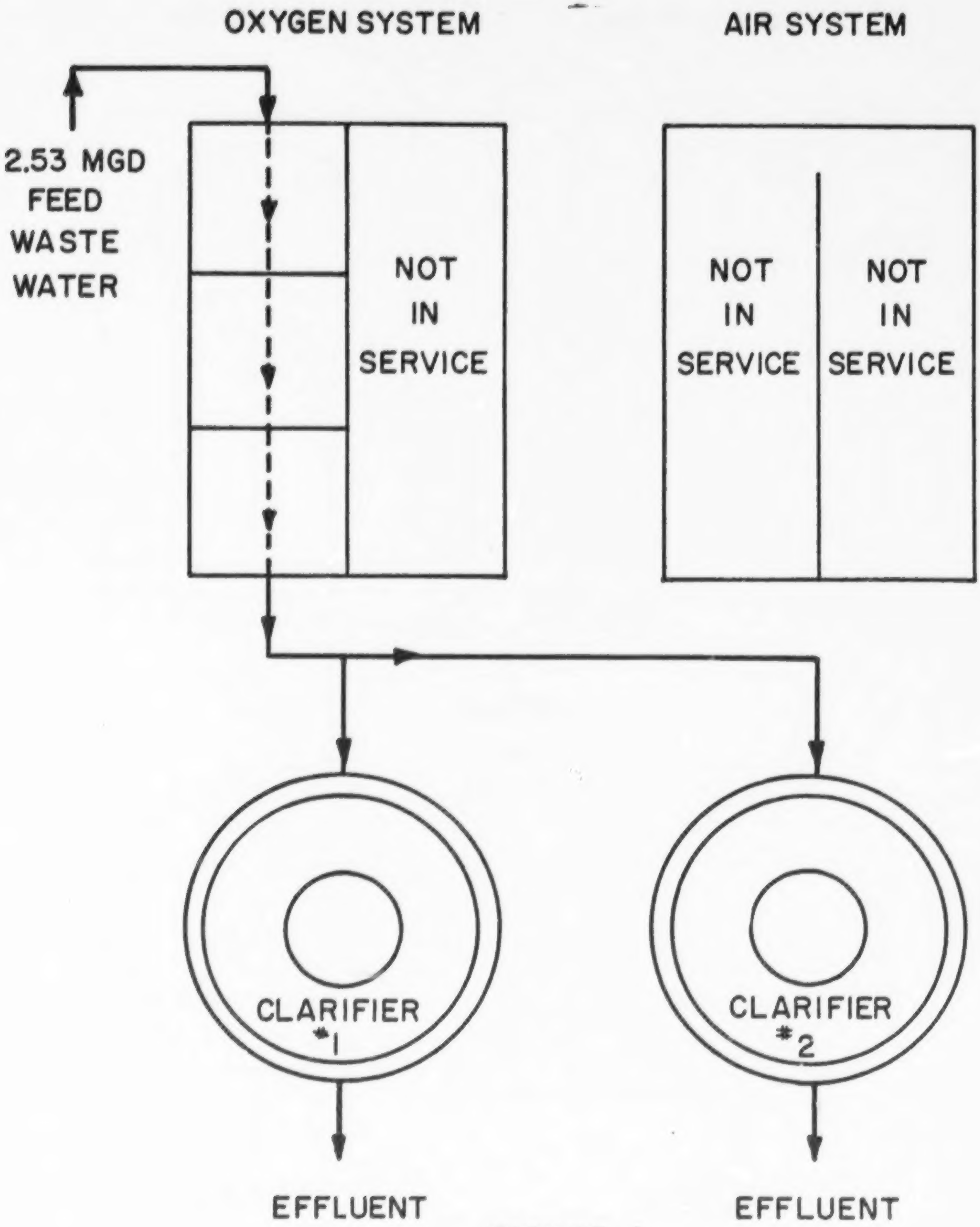

OXYGEN SYSTEM
AIR SYSTEM
2.53 MGD
FEED
WASTE
WATER
NOT
IN
SERVICE
NOT
IN
SERVICE
NOT
IN
SERVICE
CLARIFIER
#1
CLARIFIER
#2
EFFLUENT
EFFLUENT

FIGURE 4

BATAVIA TREATMENT SYSTEM
PHASE #2 OPERATION

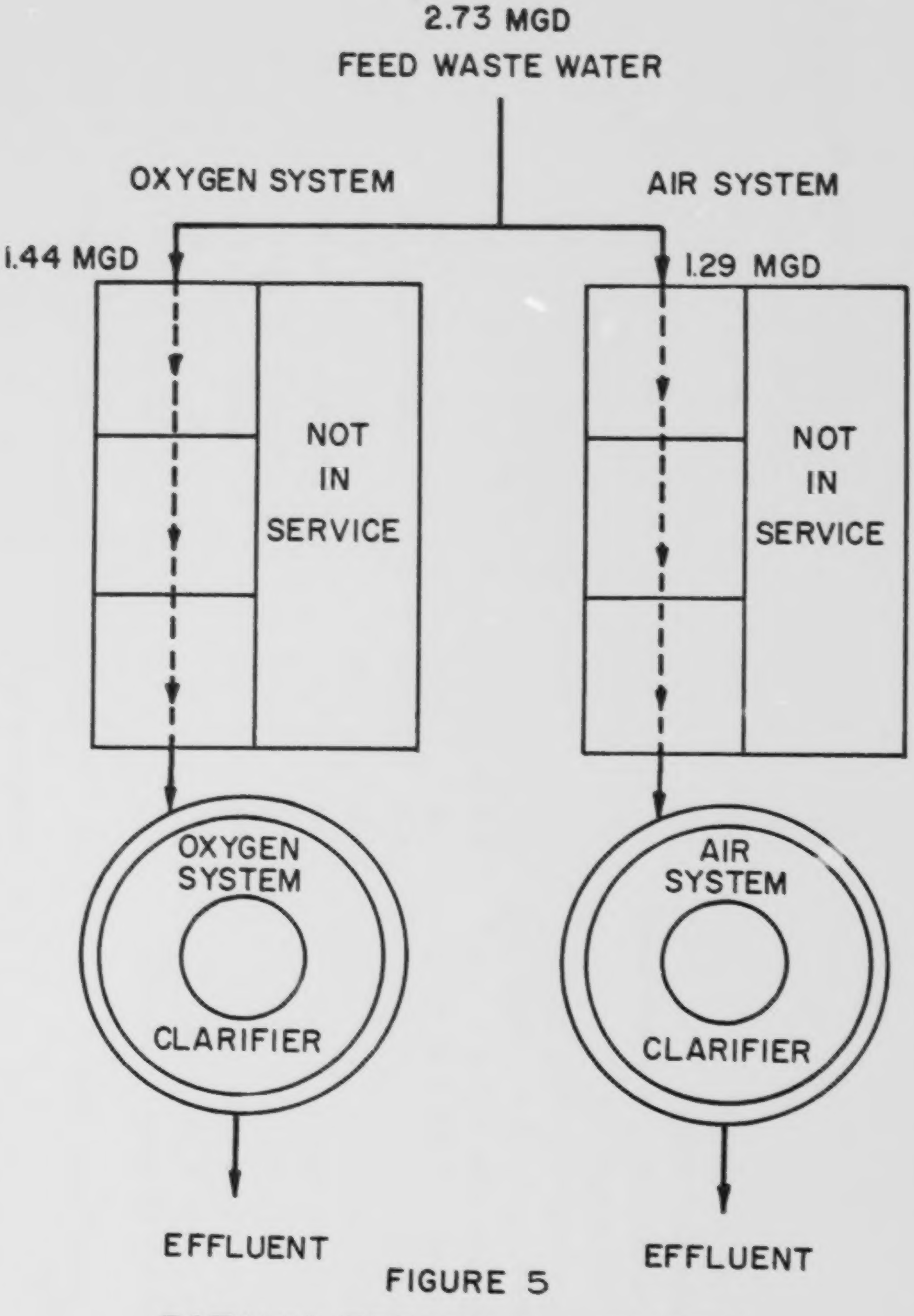

2.73 MGD
FEED WASTE WATER
OXYGEN SYSTEM
AIR SYSTEM
1.44 MGD
1.29 MGD
NOT IN SERVICE
NOT IN SERVICE
OXYGEN SYSTEM
CLARIFIER
AIR SYSTEM
CLARIFIER
EFFLUENT
EFFLUENT

FIGURE 5

BATAVIA TREATMENT SYSTEM
PHASE * 3 OPERATION

FIGURE 6

COMPARATIVE SUMMARY OF AIR AND OXYGENATION TREATMENT SYSTEM PERFORMANCE

	Phase I Operation		Phase II Operation	Phase III Operation	
	Air System	Oxygenation System	Oxygenation System	Air System	Oxygenation System
Wastewater Feed Rate (MGD)	1.97	1.91	2.53	1.29	1.44
Sludge Recycle Rate (MGD)	0.25	0.44	0.83	0.25	0.60
Recycle Sludge/Wastewater Feed Rate Ratio	0.13	0.24	0.34	0.21	0.45
Wastewater Temperature (°F)	59	59	66	70	70
Mixed-liquor average D.O. Conc.(mg/l)	1.5	8.7	9.0	0.8	8.0
Clarifier Effluent D.O. Conc. (mg/l)	0.4	4.0	0.7	0	4.3
*Aeration Detention Time (hrs.)	3.5	3.4	1.2	2.6	2.0
**Nominal Aeration Detention Time (hrs.)	4.0	4.1	1.5	3.0	2.8
MLSS Concentration (mg/l)	2440	3060	6980	3640	6190
MLVSS Concentration (mg/l)	1740	2210	4450	2580	4310
Recycle Sludge TSS (mg/l)	14,960	18,620	29,560	16,600	18,790
Recycle Sludge VSS (mg/l)	9,710	11,760	18,180	11,840	12,720
Recycle Sludge VSS/TSS Ratio	0.65	0.63	0.62	0.71	0.69
Mixed-liquor Sludge Volume Index	76	64	36	63	49
Mixed-liquor Initial Settling Velocity, (ft/hr)	7.7	7.2	6.5	7.7	5.1
Final Clarifier Overflow Rate as gal/ft^2/day	1570	1520	1010	1030	1140

* Raw flow plus recycle sludge flow.

** Raw wastewater flow only.

FIGURE 7

COMPARATIVE SUMMARY OF AIR AND OXYGENATION TREATMENT SYSTEMS PERFORMANCE

	Phase I Operation		Phase II Operation	Phase III Operation	
	Air System	Oxygenation System	Oxygenation System	Air System	Oxygenation System
Feed Wastewater BOD (mg/l)	159	159	220	262	262
Clarifier Effluent BOD (mg/l)	16	11	23	30	14
% BOD Removal	90	92	90	88	94
Feed Wastewater COD (mg/l)	352	352	325	578	578
Clarifier Effluent COD(mg/l)	84	73	97	116	89
% COD Removal	76	80	71	79	84
Feed Wastewater TSS (mg/l)	221	221	174	430	430
Clarifier Effluent TSS (mg/l)	16	9	19	23	12
% TSS Removal	93	96	89	94	97
Feed Wastewater VSS (mg/l)	152	152	123	332	332
Effluent Turbidity (J.T.U.)	5.7	5.6	9.1	5.7	4.4
Food/Biomass Ratio as lbs.BOD/Day/lb. MLVSS	0.566	0.414	0.790	0.841	0.554
Volumetric Organic Loading as lbs.BOD/day/1000 ft^3 mixed-liquor	60.0	57.9	212.5	128.9	144.8
Lbs. Dry Solids Wasted/Day	3148	1804	2448	3494	592
Lbs. VSS Wasted per Lb. BOD Removed	0.871	0.482	0.412	0.989	0.134

FIGURE 8.

COMPARATIVE SUMMARY OF AIR AND OXYGENATION TREATMENT SYSTEMS PERFORMANCE

	Phase I Operation		Phase II Operation	Phase III Operation	
	Air System	Oxygenation System	Oxygenation System	Air System	Oxygenation System
Feed Wastewater Total Phosphate(mg/1)	11	11	12	19	19
Clarifier Effluent Total Phosphate (mg/1)	6	7	10	13	14
% Total Phosphate Removed	44	39	12	32	27
Feed Wastewater NH_3-N (mg/1)	20	20	19	30	30
Clarifier Effluent NH_3-N (mg/1)	17	15	18	24	24
Feed Wastewater TKN (mg/1)	34	34	33	53	53
Clarifier Effluent TKN (mg/1)	22	19	22	31	30
Feed Wastewater NO_2-N (mg/1)	0.2	0.2	0	0	0
Clarifier Effluent NO_2-N(mg/1)	0.4	0.9	0.2	0.2	0.7
Feed Wastewater NO_3-N (mg/1)	0.3	0.3	0.1	0.1	0.1
Clarifier Effluent NO_3-N (mg/1)	0.6	1.6	1.7	2.2	2.5
Feed Wastewater Total Nitrogen (mg/1)	35	35	34	53	53
Clarifier Effluent Total Nitrogen(mg/1)	23	21	26	33	33
% Total Nitrogen Removed	35	37	23	36	36
NO_3-Nitrogen/Total Nitrogen in Clarifier Effluent	0.03	0.08	0.07	0.07	0.07
Lbs. Total Nitrogen Removed per Lb. Dry Solids Wasted	0.066	0.129	0.066	0.060	0.453

FIGURE 9.

COMPARATIVE SUMMARY OF AIR AND OXYGENATION TREATMENT SYSTEMS PERFORMANCE

	Phase I Operation		Phase II Operation	Phase III Operation	
	Air System	Oxygenation System	Oxygenation System	Air System	Oxygenation System
Ft^3 Air Utilized/gal. Sewage Treated	2.89	-	-	4.32	-
Feed Oxygen Flow Rate (cfm,NTP)	-	18.3	31.0	-	19.0
Gas Exhaust Rate from Oxygen System Final Stage (cfm,NTP)	-	1.74	4.04	-	3.22
Oxygen System Exhaust Gas Composition as % Oxygen	-	46	55	-	51
Overall % of Feed Oxygen Utilized	-	95.5	92.7	-	91.4
Lbs. Oxygen Utilized/Lb. BOD Consumed	-	0.94	0.96	-	0.72
Total Power Required for Liquid Mixing (HP)	-	26.0	11.1	-	11.7
Total Power Required for Gas Recirculation (HP)	-	2.6	9.2	-	10.6
Overall Power Utilized as HP per 1000 Gal. Mixed-liquor	-	0.088	0.125	-	0.137
Total Air Blower Power Utilized (HP)	141	-	-	138.5	-

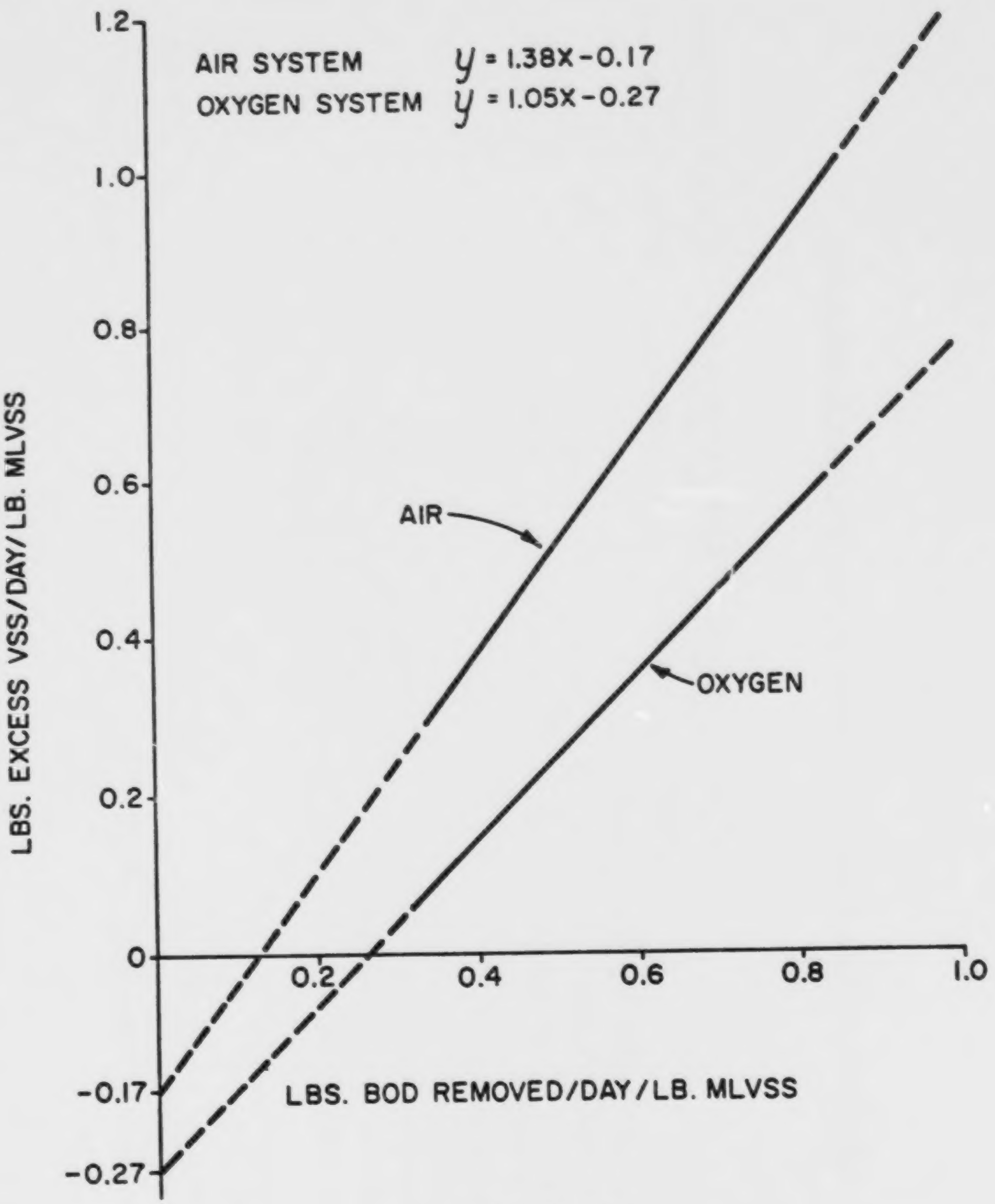


FIGURE 10

CORRELATION OF EXCESS VSS FROM
AIR AND OXYGEN SYSTEMS WITH
BOD REMOVAL AND LBS. MLVSS

CORRELATION OF % BOD REMOVAL
AT BATAVIA WITH ACTUAL DETENTION
PERIOD AND MG/L MLVSS

FIGURE 11

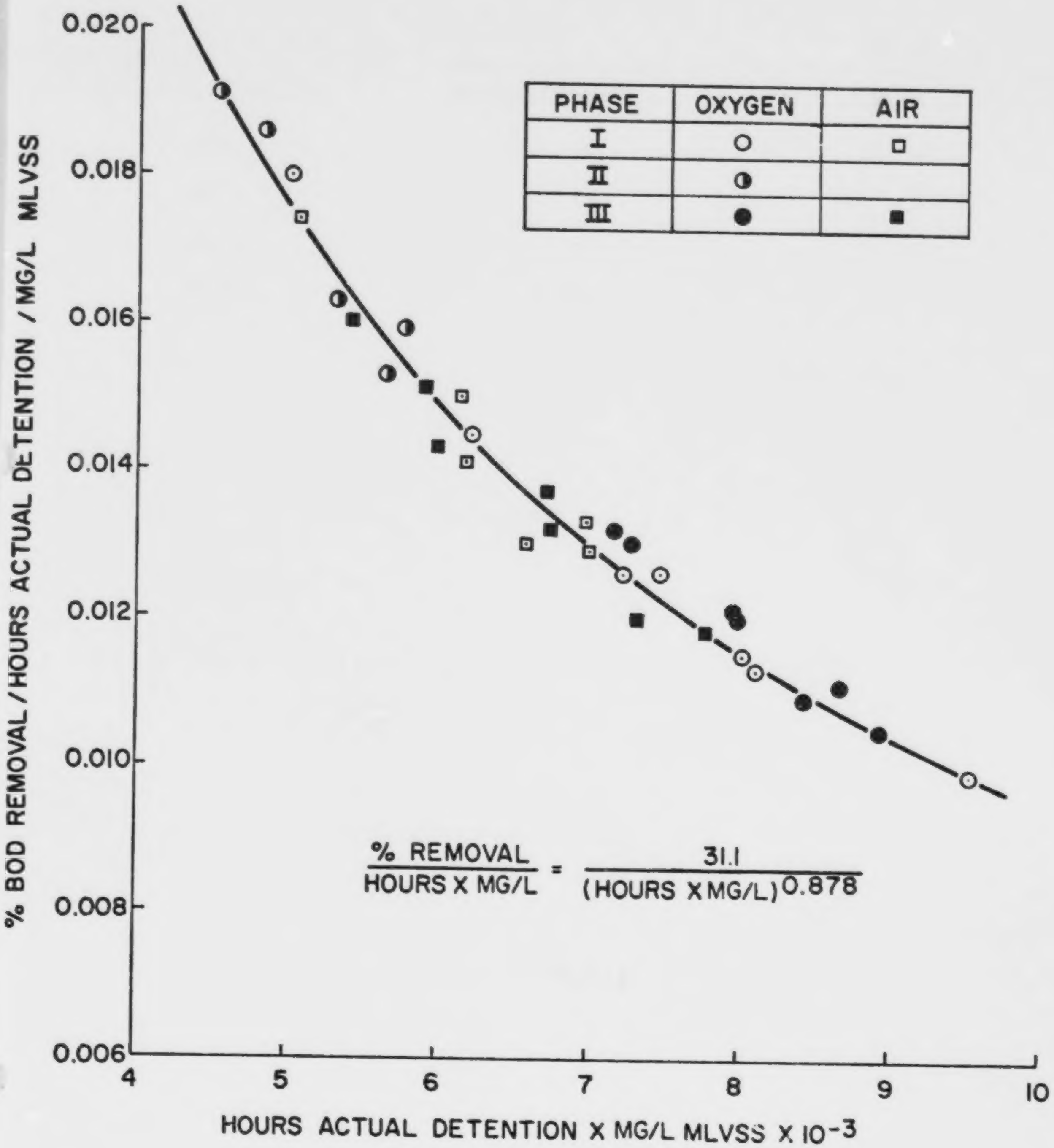


FIGURE 12

SUMMARY OF AERATION RESIDENCE TIMES AND BOD REMOVALS FOR AIR AND OXYGENATION SYSTEM ECONOMIC COMPARISONS

Air System Aeration Detention Time-Hrs. (based on Raw Flow)	Percent BOD Removal Secondary System	Oxygen System Aeration Detention Time-Hr (based on Raw Flow)
3.0	84	1.00
4.0	87	1.33
5.0	90	1.66
6.0	92	2.00

FIGURE 13.

SUMMARY OF VOLUMETRIC ORGANIC LOADINGS AND FOOD TO BIOMASS RATIOS FOR AIR AND OXYGENATION SYSTEM DESIGNS

Aeration Detention Time-Hrs(based on raw flow)	MLVSS Conc. mg/l	Volumetric Organic Loading (lbs BOD/Day 1000 ft^3)	Food to Biomass Ratio (lb BOD/Day lb MLVSS)
		OXYGENATION SYSTEM	
1.00	4420	194	.70
1.33	4420	146	.53
1.66	4420	117	.42
2.00	4420	97	.35
		AIR SYSTEM	
3.0	1625	65	.64
4.0	1625	49	.48
5.0	1625	39	.38
6.0	1625	32	.32

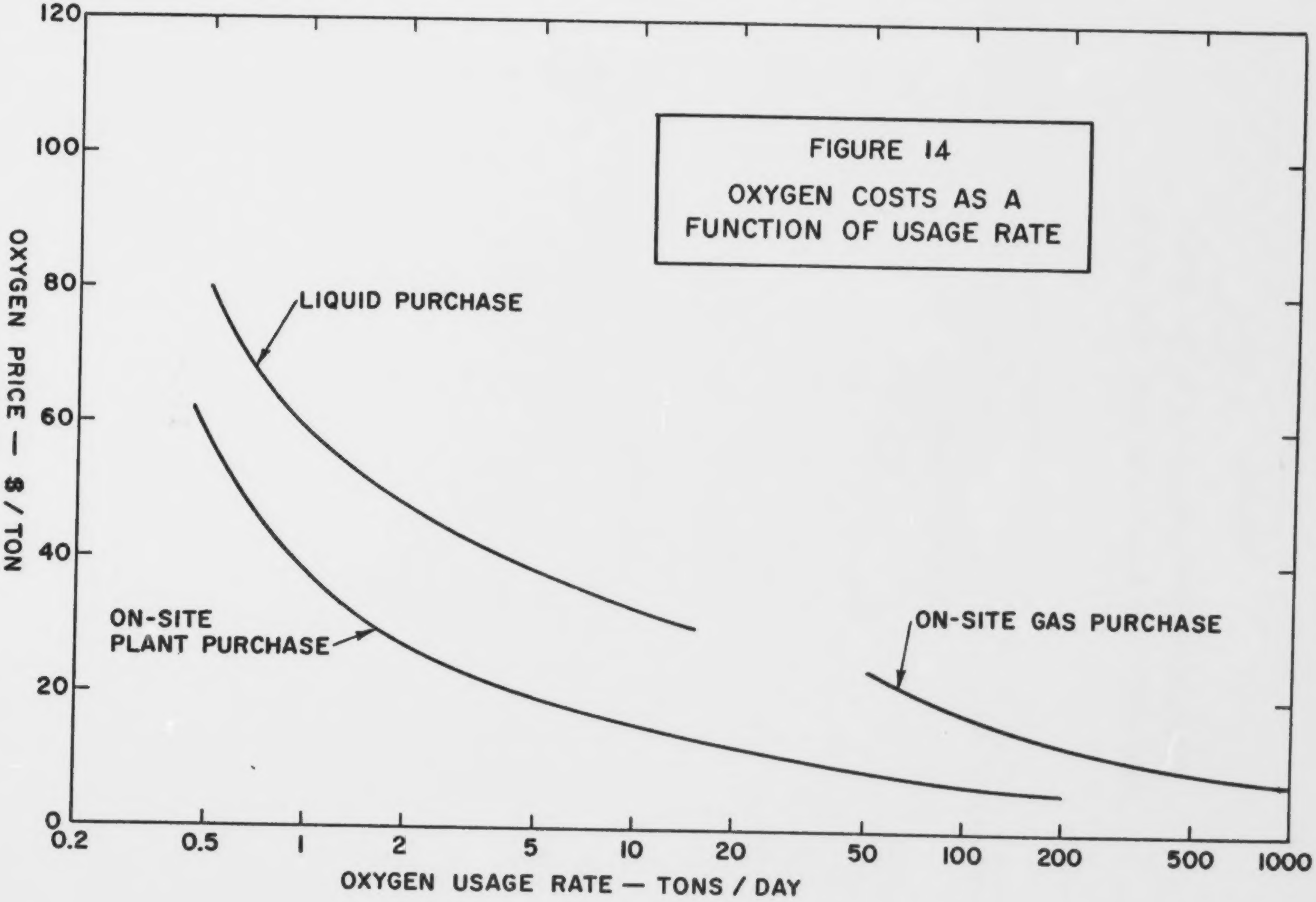


FIGURE 14

OXYGEN COSTS AS A FUNCTION OF USAGE RATE

FIGURE 15

OXYGENATION SYSTEM POWER REQUIREMENTS

MGD Treated	*Hours Detention	Oxygen Dissolution HP	Oxygen Generation HP	Total Oxygen System HP
1	1.00	6	11	17
	1.33	7	12	19
	1.66	8	14	22
	2.00	10	15	25
6	1.00	37	63	100
	1.33	43	70	113
	1.66	50	78	128
	2.00	57	85	142
30	1.00	180	300	480
	1.33	210	350	560
	1.66	250	385	635
	2.00	280	420	700
100	1.00	510	800	1310
	1.33	620	900	1520
	1.66	740	1000	1740
	2.00	870	1090	1960

* Based on Raw Sewage Flow

FIGURE 16

DIFFUSED AIR AERATION POWER REQUIREMENTS

	BLOWER HP REQUIRED		
MGD TREATED	0.8 CF (NTP) OF AIR PER GAL. OF SEWAGE	1.6 CF (NTP) OF AIR PER GAL. OF SEWAGE	2.4 CF (NTP) OF AIR PER GAL. OF SEWAGE
1	25	50	75
6	150	295	440
30	735	1470	2200
100	2450	4880	7350

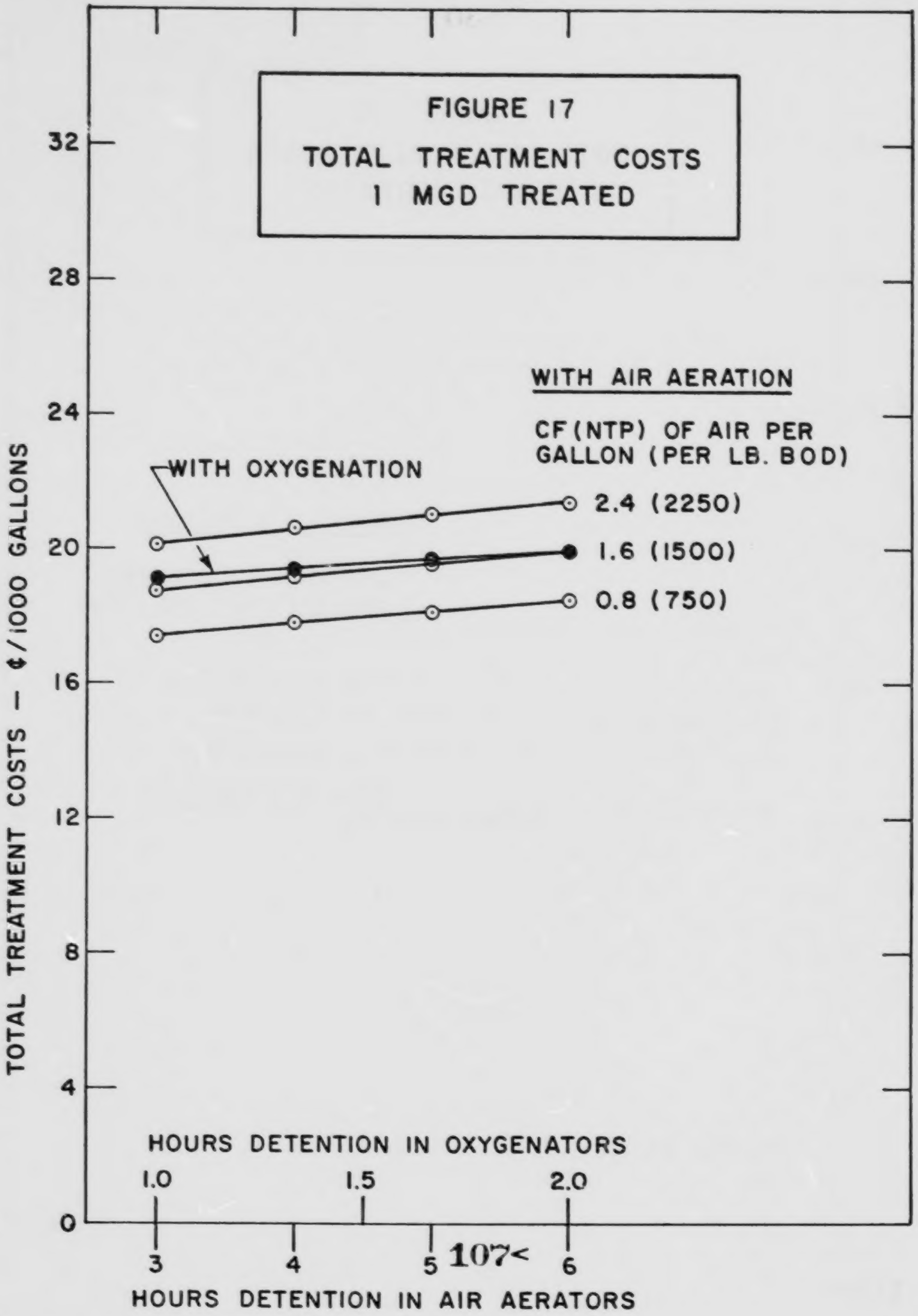


FIGURE 17

TOTAL TREATMENT COSTS
1 MGD TREATED

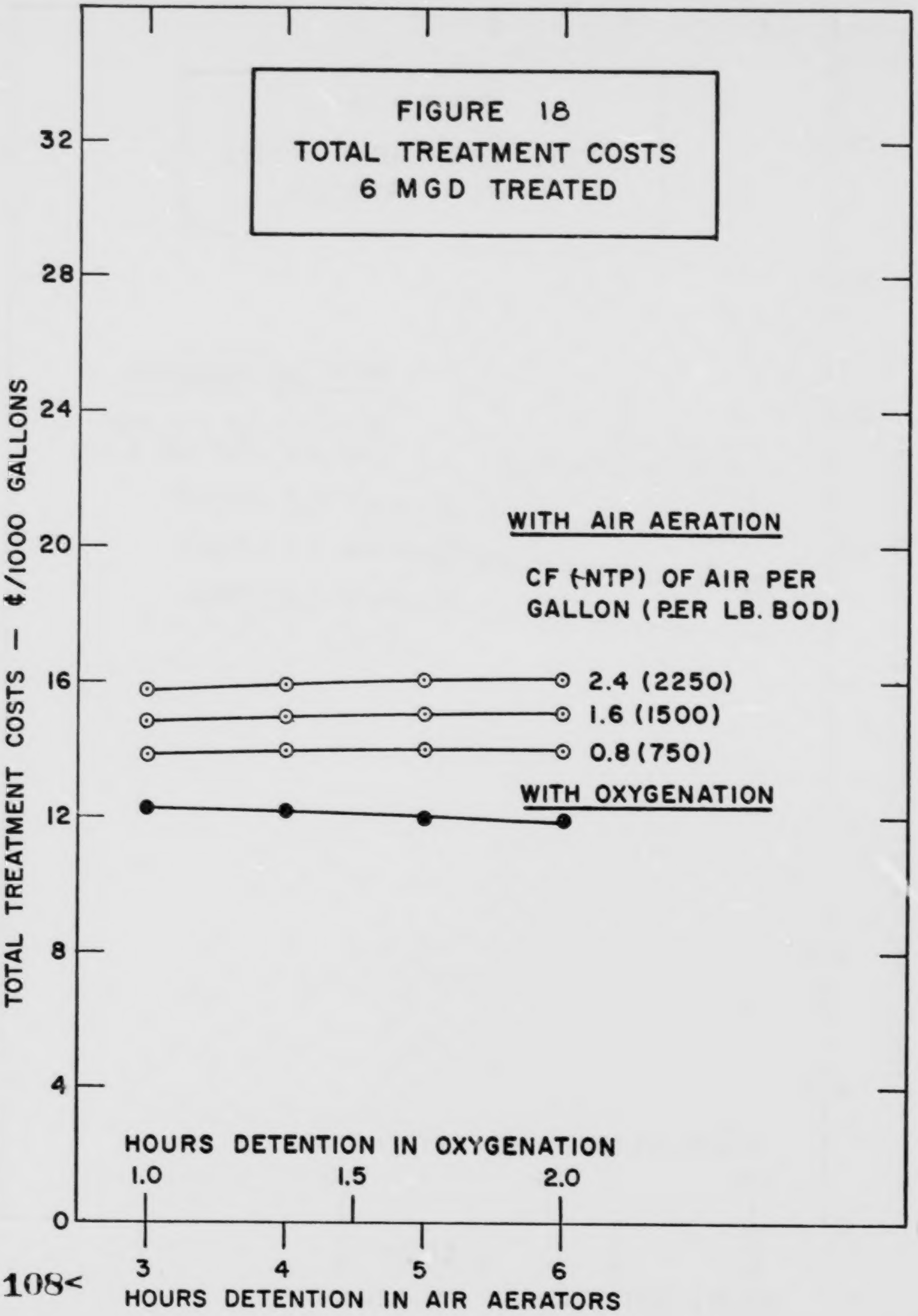


FIGURE 18
TOTAL TREATMENT COSTS
6 MGD TREATED

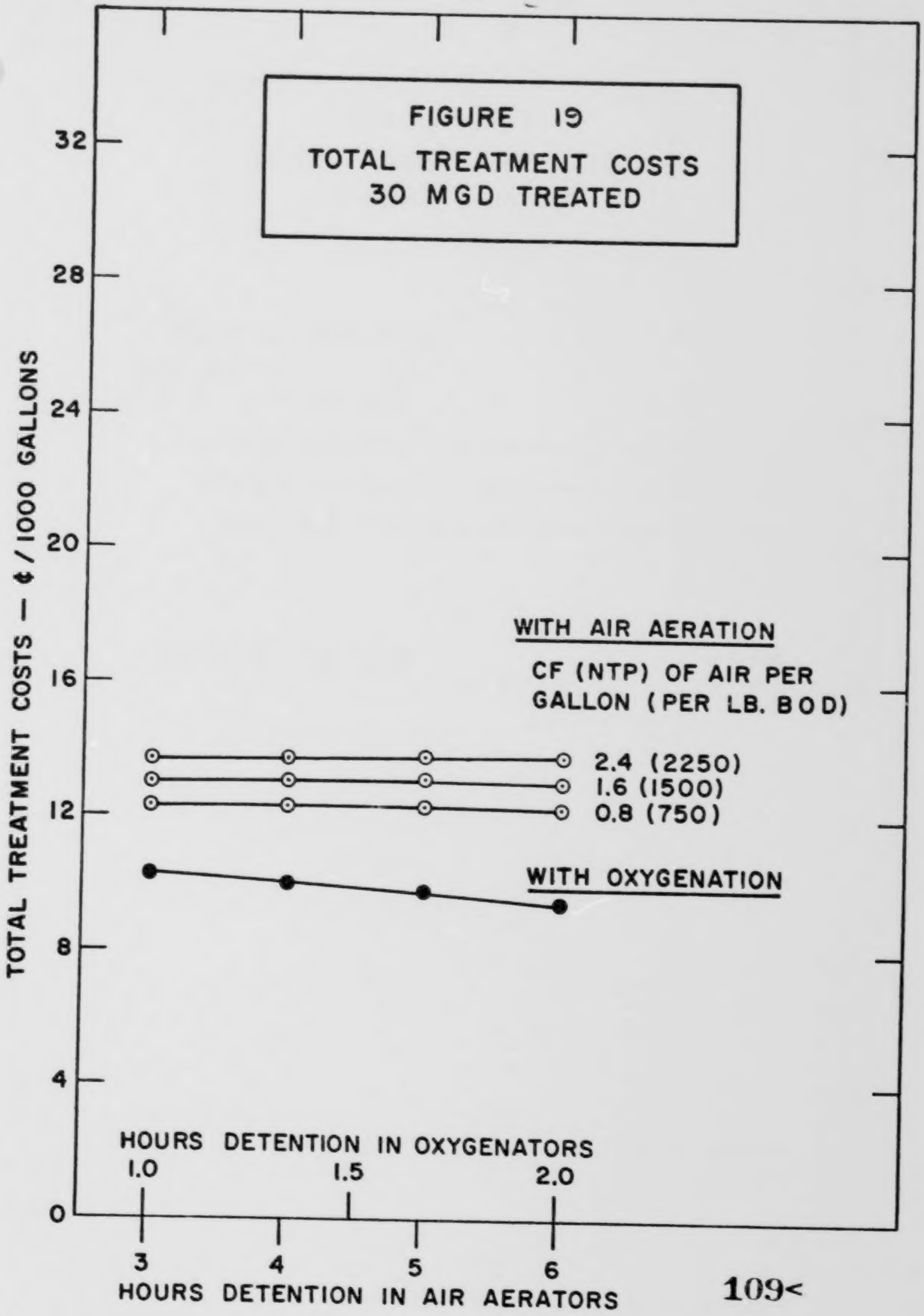


FIGURE 19

TOTAL TREATMENT COSTS
30 MGD TREATED

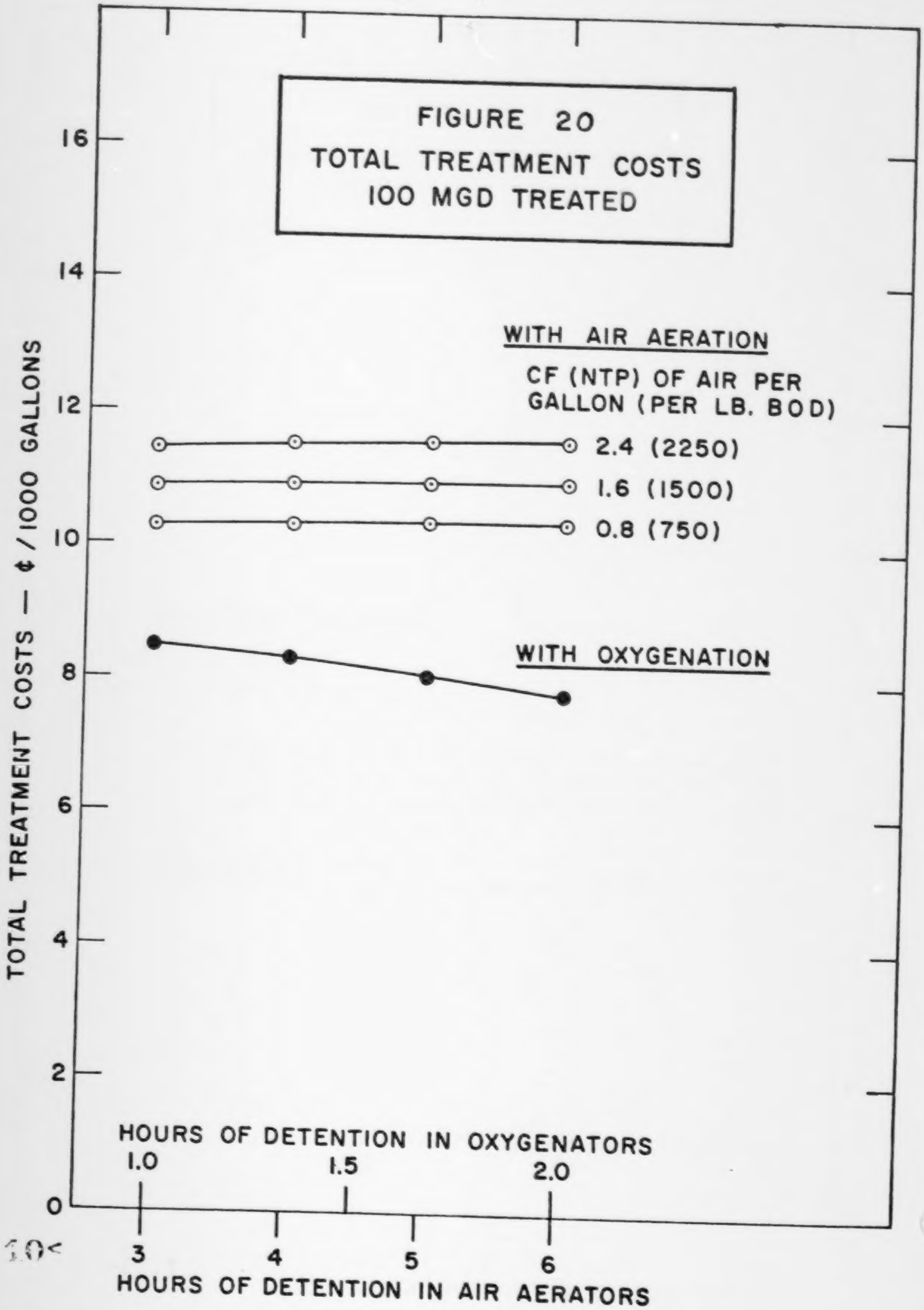


FIGURE 20
TOTAL TREATMENT COSTS
100 MGD TREATED

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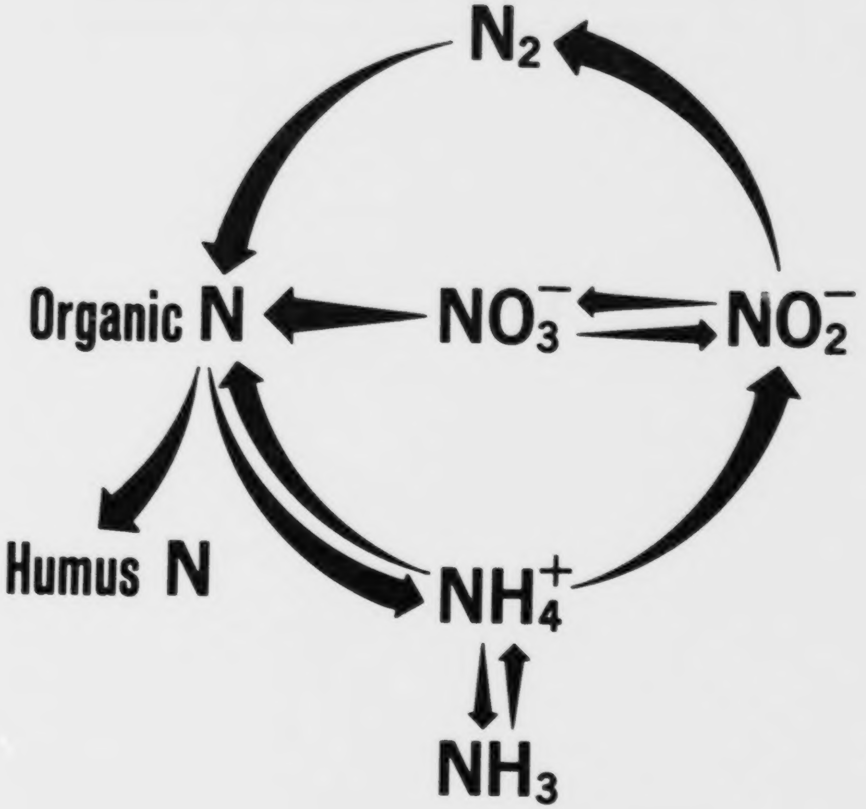
NITROGEN REMOVAL FROM WASTEWATERS


N_2
Organic N
NO_3^-
NO_2^-
Humus N
NH_4^+
NH_3

FEDERAL WATER QUALITY ADMINISTRATION
DIVISION OF RESEARCH AND DEVELOPMENT
ADVANCED WASTE TREATMENT RESEARCH LABORATORY
CINCINNATI, OHIO
OCTOBER 1970

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FOREWORD

Classically, sewage treatment has been directed toward the removal of suspended solids, biochemical oxygen demand and reduction of bacterial contaminants. Removal of nutrients such as phosphorus and nitrogen has been incidental to treatment processes. The rapid increase in severe eutrophication problems in recent years has focused attention on nutrient control from municipal sources.

Phosphorus removal techniques, as a result of extensive research over the past several years, are now useable by cities and are awaiting the extensive application that is expected. Nitrogen transformation or removal from most wastewaters will have to be accomplished. In some instances, converting ammonia to nitrates will be sufficient; in other cases, full nitrogen removal will be required.

Full scale treatment plants designed for nitrogen removal are not yet in operation in the United States. A full scale (50 mgd) plant to convert ammonia to nitrate is being designed. Biological denitrification can be designed when needed. The Advanced Waste Treatment Research Laboratory has conducted and sponsored much of the current research on nitrogen removal. This report, assembled by Dr. Robert B. Dean with the assistance of several staff members, provides a state of the knowledge on nitrogen removal that is current as of October 1, 1970. This report supersedes an earlier report dated May, 1970.

F. M. Middleton
Director of Research
Advanced Waste Treatment
 Research Laboratory

NITROGEN REMOVAL FROM WASTEWATERS

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Foreword

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4. Ammonia Nitrogen Removal by Stripping with Air — Farrell
5. Removal of Ammonia Nitrogen by Selective Ion Exchange — Dean
6. Ion Exchange for Nitrate Removal — Dobbs
7. Other Methods for Removing Nitrogen — Dean
8. Salt Creek Two Stage Nitrification Plant — Woodward

Bibliography

NITROGEN REMOVAL FROM WASTEWATERS

Edited by Robert B. Dean
Advanced Waste Treatment Research Laboratory
Robert A. Taft Water Research Center

ABSTRACT

Ammonia and nitrates are the principal forms of nitrogen in wastewaters. Ammonia is converted to nitrates by biological oxidation. Because of this reaction it exerts a high BOD in surface waters. Nitrates are reduced to nitrogen gas by bacteria when organic matter exceeds the oxygen supply. Both compounds are algal nutrients but ammonia is much more toxic than nitrates. The best developed method for control of nitrogen compounds is biological oxidation to nitrates followed by denitrification with the aid of methanol.

Ammonia stripping at high pH in cooling towers following lime treatment is effective but cannot be used during freezing weather and may suffer from serious scale problems. Selective ion exchange of ammonia with lime regeneration may be practical but the process is still in the pilot stage. Several other processes are being studied. A bibliography of pertinent references is attached.

Paper No. 1

NITROGEN REMOVAL FROM WASTEWATERS -- STATEMENT OF THE PROBLEM

Edwin F. Barth and Robert B. Dean

Nitrogen is continuously cycled through the environment from the atmosphere to growing plants and back to nitrogen gas. The principal species are presented in the classical nitrogen cycle as shown on the cover. Fixation of nitrogen to produce organic nitrogen is carried out by specialized bacteria, algae and a few higher plants. When these organisms decompose much of the organic nitrogen is converted to ammonia. A small fraction however is converted to relatively stable humus nitrogen in the soil. Ammonia and ammonium ion are rapidly interconvertible and the ratio of the two species depends on the pH according to the equation:

1) $\log (NH_3/NH_4) = pH - 9.25$ at 25°

Minor corrections to this equation are necessary in brackish waters and at low or high temperatures. Below pH 7.25 less than 1% of ammonia is present as NH_3.

Ammonia is nitrified or converted to nitrates by two specialized groups of bacteria. *Nitrosomonas* oxidizes ammonia to nitrite and *Nitrobacter* oxidizes nitrite to nitrate. Oxygen is required as well as CO_2 for bacterial growth but dissolved organic matter has little or no influence. Nitrifiers are very sensitive to poisoning by many simple substances including heavy metals and free ammonia. Biological nitrification of industrial wastes can therefore not be assumed without careful testing under realistic conditions.

Nitrates can be reduced by many bacteria that use the oxygen for metabolism. Nitrites are the first product but they are usually rapidly reduced to nitrogen gas by further bacterial action. Nitrates can also be utilized as a source of nitrogen by most growing plants.

The nitrogen content of natural unpolluted waters is normally less than one milligram per liter, and during the growing season, soluble nitrogen compounds are virtually completely depleted by growing plants and algae (AWWA, 1970).* The presence of nitrogen compounds, particularly ammonia, has therefore been taken as an indication of pollution for over a century. Historically, nitrogen analyses have seniority over BOD$_5$ and suspended solids determinations to trace wastewater pollution and to control biological treatment. The BOD$_5$ tests replaced tests for various nitrogen compounds for two reasons. First, analysis was not as tedious; and secondly, the BOD$_5$ tests made biological processes look more efficient, since it is much easier to get 85% reduction of BOD than it is to get a comparable conversion to nitrates.

Municipal wastewaters have nitrogen contents in the 15-25 mg/l range in untreated and primary settled wastes; the nitrogen is divided between organic compounds, which are mostly insoluble, and ammonia. Biological treatment converts most of the organic nitrogen to ammonia, and under favorable conditions ammonia may be oxidized to nitrates. Nitrogen compounds of lesser significance are discussed in paper No. 7.

The conditions for nitrification of municipal effluents have been worked out by Downing and Hopwood (1964) who showed that

> "to achieve nitrification consistently, the period of aeration must exceed a minimum value which is a function of the concentration of activated sludge, the temperature, and the strength of the sewage. When conditions are favourable for nitrification, the rate of nitrification and the rate of consumption of oxygen due to this process will tend to an equilibrium level which is proportional to the concentration of activated sludge under aeration. In an aeration unit through which the liquid travels with ideal piston-flow, the concentration of ammonia in the effluent under the above conditions will be normally very small. The effect of longitudinal mixing will be to increase slightly the concentration of ammonia in the effluent."

---

* All references are collected in the bibliography.

Biological oxidation of ammonia consumes oxygen and, by definition in Standard Methods, is included as part of the BOD_5 test. However, much of the ammonia demand may not show up in five days because of inadequate seed bacteria or excessive competition from heterotrophic forms. The BOD_{20} is more realistic in assessing the oxygen demand due to ammonia in municipal effluents. The oxygen demand from ammonia in industrial effluents is best calculated assuming complete nitrification.

1) $2NH_3 + 4O_2 \longrightarrow 2HNO_3 + 2H_2O$

According to this equation, 1 mg of ammonia nitrogen could exert an ultimate BOD of 4.5 mg of oxygen. Well over 60% of the biological oxygen demand that will be exerted in receiving waters by biologically treated municipal effluents will be due to oxidation of ammonia to nitrate by the stream's flora. The BOD bottle is a relatively inefficient system for nitrification in comparison with the attached flora on the beds of the stream; therefore, reliance on the 5-day BOD bottle has shifted the burden of satisfying nitrogenous oxygen demand from the treatment plant to the receiving water.

It would be beneficial if waste treatment plants were required to produce nitrified effluent. Ammonia nitrogen in effluent has several undesirable features:

(1) Ammonia consumes dissolved oxygen in the receiving water;
(2) Ammonia reacts with chlorine to form chloramines which are less effective disinfectants than free chlorine;
(3) Ammonia is toxic to fish life;
(4) Ammonia is corrosive to copper fittings;
(5) Ammonia increases the chlorine demand at waterworks downstream.

A nitrified effluent, free of substantial concentrations of ammonia, offers several advantages:

(1) Nitrates will provide oxygen to sludge beds and prevent the formation of septic odors;

(2) Nitrified effluents are more effectively and efficiently disinfected by chlorine treatment;

(3) A nitrified effluent contains less soluble organic matter than the same effluent before nitrification.

In general, we can depend on conventional biological processes to transform almost all nitrogenous components in wastewater into ammonia and biological sludge. Once this has been accomplished, we can design systems to remove ammonia by air-stripping, or we can proceed to a second biological conversion of the ammonia into nitrates. A nitrified effluent may be acceptable for discharge, but if the nitrate level is too high it can be further treated by biological action to convert the nitrates to nitrogen gas.

Ammonia and nitrate are interchangeable nitrogenous nutrients for green plants and algae, as well as bacteria. Some species may prefer ammonia, and others nitrate, but in general they can be considered interchangeable (Flaigg and Reif, 1954). When algae grown on nitrates decompose they will liberate nitrogen in the form of ammonia.

Nitrogen and Eutrophication

The head of the National Eutrophical Program of the FWPCA has stated (Bartsch, 1970),

> "In many lakes that are in trouble because of eutrophication, the nutrients of prime significance are nitrogen and phosphorus. Phosphorus is found to be critical in terms of extensive production more frequently than nitrogen. In some cases it has been demonstrated through field study and appropriate assays that nitrogen does occupy a critical role. Reasons for pessimism relative to nitrogen control (of eutrophication) are the following:
>
> "(1) Many microorganisms can fix atmospheric nitrogen and make it available. Included are bacteria, some of the blue-green algae which are causes of bloom problems, and actinomycetes.

> "(2) Nitrogen can pass through the soil and into the ground water and find its way into surface waters more easily than phosphorus. Ammonia can also find its way into surface water via precipitation, in some locations up to as much as 20 lbs/acre per year.
>
> "Situations where nitrogen control must be used for control of eutrophication include the upper San Francisco Bay, where it has been demonstrated that nitrogen is the critical element. There is some thought that lakes that are young in the aging process are more likely to be nitrogen limiting than lakes that have started down the pathway of eutrophication. Lake Tahoe is an example where nitrogen may or may not be a limiting nutrient. In some bodies of water where phosphorus may be limiting part of the time during the growing season, nitrogen may normally be limiting at other times. It is possible to exercise some guided judgment in the matter of limiting nutrients through use of test procedures like the Provisional Algal Assay Procedure (PAAP, 1969)."

Nitrogenous Oxygen Demand

The Chief of the Engineering Section, Chesapeake Technical Support Laboratory, Middle Atlantic Region, FWQA, has stated in regard to nitrogen removal needs (Jaworski, 1970):

> "In the upper Potomac Estuary about one half of the dissolved oxygen sag is due to a nitrogenous demand of organic matter being discharged from wastewater effluents in the Washington metropolitan area. During the low-flow summer months, dissolved oxygen levels of less than 2 mg/l in a 10-mile reach are common. Studies by CTSL, FWQA and Maryland Department of Water Resources have established that the dissolved oxygen deficits in the middle portion of the Patuxent River basin are mainly attributable to the nitrogenous demand of organic matter originating from wastewater discharges.
>
> "To eliminate the dissolved oxygen deficit in the Potomac Estuary and in the Patuxent River, the removal of nitrogenous demand is required to manage the water quality in both areas. The State of Maryland has gone on record as requiring a maximum of 1 mg/l of ammonia in effluents discharging into the Patuxent River. To enhance the water quality of the Potomac River basin, the recent Potomac Enforcement Conference required 85% removal of the present nitrogenous oxygen demand from treated waste discharges in the Washington metropolitan area.

Algal Growth Response

> "At the present time, predictive generalizations cannot be made for the response of algae to nutrients for all receiving waters. Different geophysical systems appear to be responsive to different limiting nutrients.

> "In the Potomac Estuary during summer periods, nitrogen appears to be the rate limiting nutrient at the salinity-freshwater interface. Data from the James River Estuary by VIMS also indicate that the rate limiting nutrient is nitrogen. Thus it appears that nitrogen can be a manageable nutrient with respect to the control of eutrophication.
>
> "If a wastewater system is selected that converts all or most of the nitrogen to nitrates, this will aid the immediate oxygen deficit caused by the oxidation of NH_3 to NO_3. Since NO_3 can be readily converted to organic nitrogen and NH_3 by bacteria and algal cells in waters, the potential nitrogenous demand will still be present. Therefore, carbon and phosphorus as well as nitrogen are the important parameters in the control of eutrophication with respect to the elimination of nuisance algal conditions and the enhancement of water quality.
>
> "While the control of organic carbon and phosphorus will reduce the incidence of algal blooms, the addition of nitrogen removal will further enhance the control of the blooms. If wastewater treatment processes are added to convert the nitrogen to NO_3 without considering the complete removal of nitrogen, we are just transporting the problem farther downstream."

The State of Illinois has recently set an effluent standard of 2.5 mg/l of ammonia (2.05 mg/l ammonia-N) for discharge to surface waters. A substantial fraction of the oxygen sag in the Grand and Clinton Rivers in Michigan and in the Kanawha River, West Virginia, below a large industrial complex, has been attributed to the BOD of ammonia. It was early practice in Britain to nitrify effluents as evidence of good biological treatment since nitrification is difficult in the presence of large amounts of readily decomposable organic matter. Furthermore, nitrates in the water provide oxygen to help oxidize organic matter present from other sources, and to prevent septicity. Nitrates, usually as chili saltpeter, have frequently been used to control septic odors in estuaries and lagoons (Lang, 1966).

A review article of British water pollution control (Jenkins, 1970) states that a recent requirement limits ammonia nitrogen in sewage effluents not to exceed 10 mg/l. This means that a treatment plant must be designed to give complete purification to three times the dry weather sewage flow and that 60% to 70% of the ammonia nitrogen must be oxidized at all times. River control agencies regard nitrification as desirable because of the high oxygen demand of ammonia nitrogen, the toxicity of ammonia to fish under certain circumstances, the reduction in chlorine demand, and the oxygen reserve present in nitrates.

Ammonia standards for several states were reviewed by Battelle Northwest (1969).

Ammonia is rapidly adsorbed by soil minerals and particulate matter containing nitrogen is also effectively removed in the soil. However, if there is not sufficient plant growth in the soil to use the bound ammonia, it will be converted to nitrates by nitrifying bacteria. [illegible]trates are very poorly adsorbed in the soil, and rapidly reach the ground water when present in excess. Therefore, over-fertilization with nitrogen compounds, whether they be ammonia, slow-release nitrogen, or nitrates, can lead to high concentrations of nitrates in ground water.

Ammonia has very little toxicity to humans and there are no limits on ammonia in the USPHS Drinking Water Standards of 1962, or the WHO International Standards. Ammonium compounds have been used in cooking and ammonia gas is the essential ingredient of smelling salts. Various suggested limits on ammonia in drinking water are related to the fact that ammonia is generally taken as an indicator of fecal pollution and the fact that high levels of ammonia dissolve copper from plumbing. The toxicity of ammonia to fish and other aquatic species is much higher than for humans, but details have not been completely elucidated. According to many references, the toxicity of ammonia and ammonium salts to aquatic animals is directly related to the amount of undissociated ammonium hydroxide in the solution, which in turn is a function of pH. Toxicity has been reported at levels as low as 0.2 mg/l of undissociated ammonia nitrogen (McKee and Wolf, 1963).

Nitrates have low toxicity for plants and animals. Saltpeter (potassium nitrate) has traditionally been used as a curing agent for ham. Nitrates in food and water are normally denitrified by bacteria in the intestines. However, some very young babies do not have a complete intestinal flora. The bacteria which are present reduce nitrates to nitrites, and do not carry the reduction further to nitrogen gas. Nitrites combine with hemoglobin to produce methemoglobinemia, a disease characterized by cyanosis. PHS and WHO standards recommend not more than 10 mg/l of nitrate nitrogen (45 mg/l nitrate) for water to be fed to babies. Many well waters contain more than 10 mg/l of nitrate nitrogen, and particularly among poorer classes and migrant laborers it is very difficult to make sure that babies are not given this water.

The following papers discuss specific processes which can be used to remove nitrogen compounds from water.

Paper No. 2

NITROGEN REMOVAL BY BIOLOGICAL SUSPENDED GROWTH REACTORS

E. F. Barth

Other sections of this report deal with chemical and physical processes to remove nitrogenous components of wastewater. While these are being actively investigated, our basic capability resides in carefully managed biological systems. Success in providing a high efficiency for nitrogen removal by biological denitrification requires that the biological transformation of ammonia nitrogen to nitrate nitrogen be under good process control. Any reduced nitrogen compounds introduced into the denitrification stage will pass through the process unaltered and impair overall nitrogen removal efficiency.

Downing and Hopwood (1964) revealed the complex factors involved in maintaining nitrification with a conventional activated sludge system. A three sludge variation of the activated sludge process has been developed which greatly simplifies the process control problems associated with maintaining nitrification (Barth, et al., 1968). Figure 1 is a schematic of the process developed for nitrogen removal by biological denitrification. Figure 2 is a representation of a municipal application of the three sludge system taking advantage of common wall construction techniques for economy in construction.

The three sludge system allows management of the separate biological transformations which are necessary for successful denitrification. As shown on Figure 1 the high rate system handles the bulk of the carbonaceous removal and at this station the waste activated sludge is removed. Thus, the nitrification stage receives a predominantly ammonia nitrogen feed and an enriched culture develops because each system has its own sludge recycle. This process design also has other desirable features. The high rate system protects subsequent nitrification stages from toxic chemicals. Heavy metals, cyanides, thiocyanates, and toxic organic chemicals will be either sorbed or biologically degraded before they reach the nitrification stage. Since this is a staged system there can be no direct short circuiting of materials from the influent to the effluent. Temperature effects on the enriched culture of the nitrification stage are not as extreme as with a single sludge system which contains only a marginal population of nitrifying organisms.

Once controlled nitrification has been established the biological denitrification process can be optimized. The nitrified effluent flows to a stirred anaerobic reactor where methyl alcohol is added in proportion to the nitrate nitrogen concentration. The organisms in this stage use the oxygen component of the nitrate radical to oxidize the organic carbon of methyl alcohol. The end products of this metabolism are elemental inert nitrogen gas and carbon dioxide, which are liberated to the atmosphere.

The stage approach to nitrification has been investigated through grant and contract activities and work at the Robert A. Taft Water Research Laboratory. Large pilot plant operations are located at the University of Notre Dame and Manassas, Virginia. The process has also been evaluated at a 1 MGD scale at Hazel Crest, Illinois. Temperatures as low as 6° C have been recorded, with no detrimental effect on nitrification. Previous experience with the stage approach to nitrification has been reported by English and South African investigators (Osborn, 1965).

Biological denitrification has been studied by microbiologists for many years, particularly in regard to loss of nitrogen from soils (van Schreven, 1963). There have been intensive recent studies related to the application of denitrification to wastewaters. Wuhrman (1960) and Bringmann and Kuhn (1965) have both published in Europe. American experience includes studies at the Robert A. Taft Water Research Laboratory (Barth, et al., 1968); the Lebanon Plant, Firebaugh, California (McCarty, et al., 1969); Notre Dame, Indiana (Echelberger and Tenney, 1969); Manassas, Virginia (Mulbarger, et al., 1969); and Pomona, California (Carey, 1968). A summation of these studies show that biological denitrification is a controllable process if the reaction is forced with an organic supplement, such as methyl alcohol. Attempts to use waste organic matter from sewage have been less successful.

Packed column processes as well as suspended growth reactors have given efficient nitrogen removal. Large scale municipal application, in the range of 2 MGD or greater would favor suspended growth reactors. The 300 MGD Blue Plains Washington, D. C. municipal facility tentatively is being designed on the basis of a three-stage biological system for both nitrogen and phosphorus removal after an extensive review of both chemical-physical approaches and biological approaches to nitrogen removal.

The Blue Plains plant will employ a final filtration step to assure quality control of the effluent. Filtration of effluents is necessary for high efficiency whether we are considering suspended solids, chemical oxygen demand, nitrogen, or phosphorus. We must have positive control of the solids no matter what treatment process we employ.

Figure 2 showing the municipal application of this process indicates that the current technology development of high purity oxygen for carbonaceous removal can be combined with this staged system to give very reasonable detention times. A two hour oxygen aeration period combined with one hour detention in each of the settlers, and two hours each for nitrification and denitrification give a total flow through time of ten hours.

The three-stage biological system can produce an effluent containing 2 mg/l, or less, of total nitrogen. This is a very realistic figure to use as an effluent standard, and when dealing with municipal wastewaters it represents 90 percent or greater removal of total nitrogenous materials.

The capital cost of such a system would be about 25 percent more than a conventionally designed activated sludge process of equal flow. The use of high purity oxygen in the first and second stages could bring the capital cost almost exactly in line with conventional capital costs.

The operating cost for the methyl alcohol to supplement the denitrification process is 1.6 cents per thousand gallons for a wastewater containing 20 mg/l of nitrogenous material.

In order to make more information available than is possible in the above narrative summary, reproductions of the slides used to present this topic, at Dallas, Texas, are attached after Figure 2.

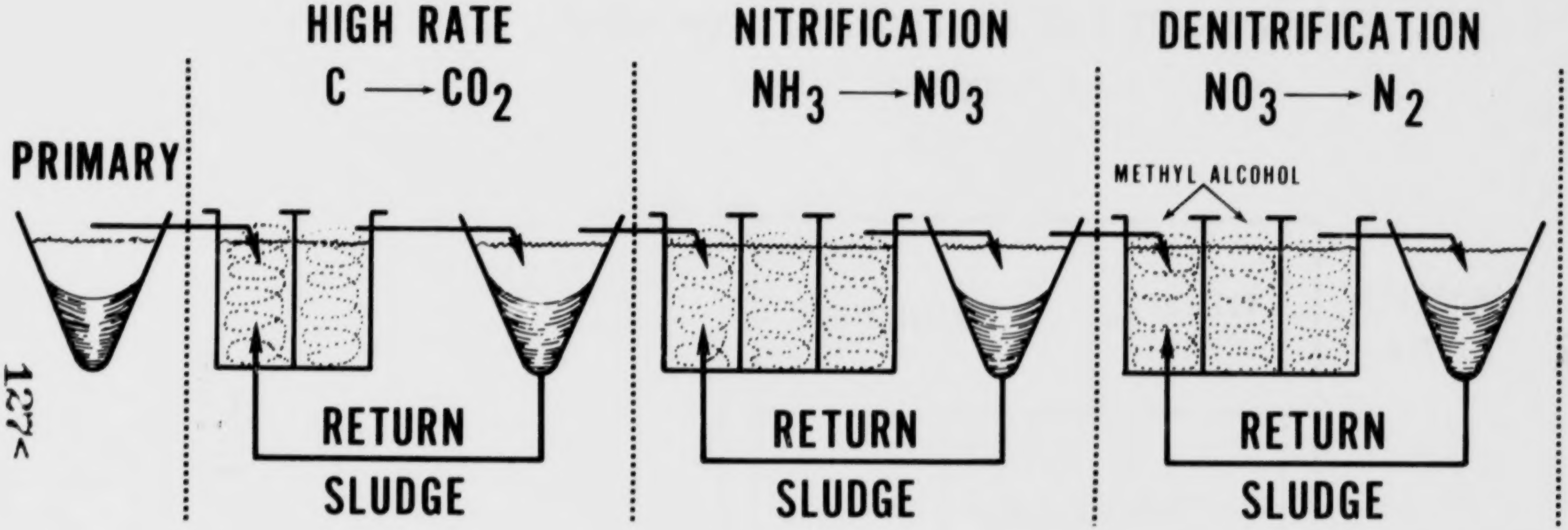

HIGH RATE
$C \longrightarrow CO_2$
NITRIFICATION
$NH_3 \longrightarrow NO_3$
DENITRIFICATION
$NO_3 \longrightarrow N_2$
PRIMARY
METHYL ALCOHOL
RETURN SLUDGE
RETURN SLUDGE
RETURN SLUDGE

FIGURE I

THREE SLUDGE SYSTEM FOR NITROGEN REMOVAL

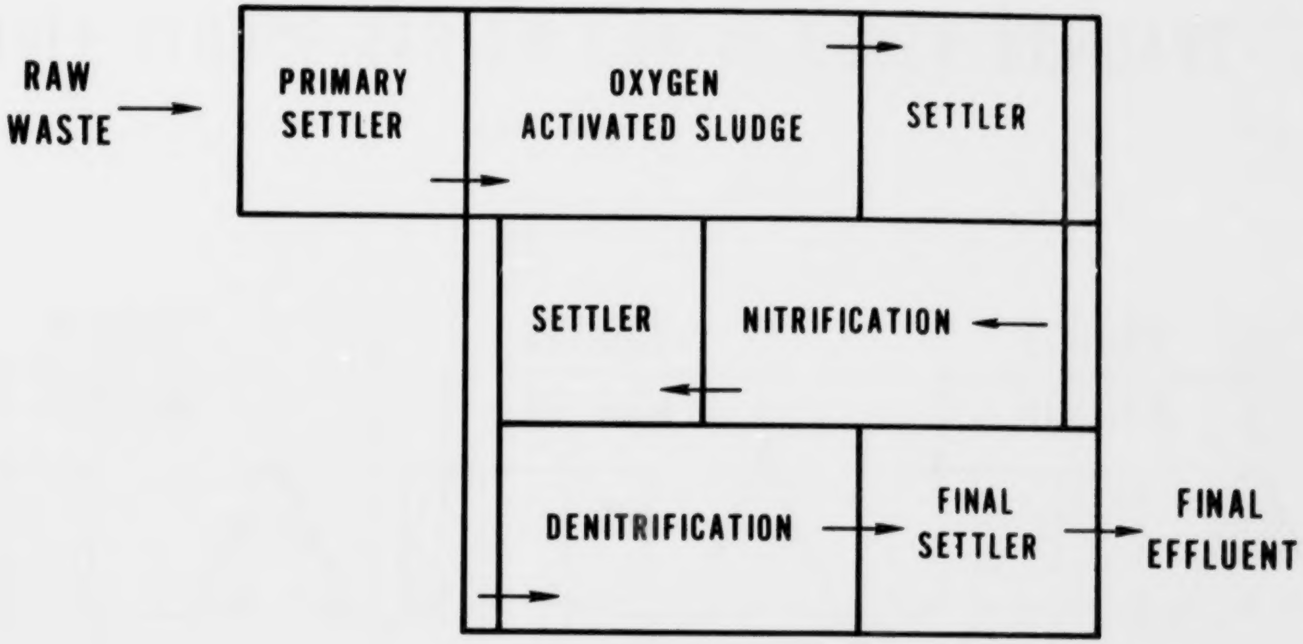

RAW
WASTE
PRIMARY
SETTLER
OXYGEN
ACTIVATED SLUDGE
SETTLER
SETTLER
NITRIFICATION
DENITRIFICATION
FINAL
SETTLER
FINAL
EFFLUENT

FIGURE 2

COMMON WALL CONSTRUCTION OF 3-SLUDGE SYSTEM

GPO 821-871-9

TABLE II

IMPORTANCE OF NITROGEN

NH_3 IN EFFLUENTS CAN CAUSE DO SAG IN RECEIVING WATER

NH_3 IS CORROSIVE TO COPPER FITTINGS

1 NH_3 REQUIRES 7 PLUS $C\ell_2$ FOR BREAKPOINT

NO_2 CAUSES HIGH $C\ell_2$ DEMAND

NH_3 INFLUENCES $C\ell_2$ CONTACT TIME

NITROGEN COMPOUNDS ARE NUTRIENTS

NO_3 CAN BE HEALTH HAZARD

NH_3 CAN BE TOXIC TO FISH

TABLE III

RELATIONSHIP OF CARBONACEOUS OXIDATION AND AMMONIA OXIDATION, BEFORE AND AFTER TREATMENT TO REMOVE 90% OF CARBONACEOUS MATERIAL

	mg/l	
	WASTEWATER	FINAL EFFLUENT
ORGANIC MATTER	250	25
ORGANIC OXYGEN DEMAND	375	37
NH_3	25	20
AMMONIA OXYGEN DEMAND	112	90
TOTAL OXYGEN DEMAND	487	127
PERCENT OF OXYGEN DEMAND DUE TO NH_3	22%	71%

TABLE IV

NITROGEN REMOVAL PROCESSES

$$NH_3^\circ \text{ (SOLUTION) } \xrightarrow{\text{air}} NH_3^\circ\uparrow \text{ (GAS)}$$

$$NH_4^+ \longrightarrow \text{CLINOPTILOLITE} \bullet NH_4$$

$$NO_3^- \longrightarrow \text{RESIN} \bullet NO_3$$

REVERSE OSMOSIS

ELECTRODIALYSIS

DISTILLATION

$$2\ NH_3 + 3\ Cl_2 \longrightarrow N_2^\circ\uparrow + 6\ HCl$$

$$NO_3^- + Fe^{++} \xrightarrow{\text{Cu}} N_2^\circ\uparrow + Fe^{+++}$$

$$NO_3^- \xrightarrow[\text{organisms}]{\text{organic C}} N_2^\circ\uparrow + CO_2$$

UNIT EFFICIENCIES, PER CENT REMOVAL

Plant	Primary			Secondary			Overall		
	COD	SS	N	COD	SS	N	COD	SS	N
Archbold	43	58	9	89	87	57	94	94	61
Hamilton	55	57	-27	77	86	34	90	94	16
Lebanon	62	65	28	45	55	42	79	84	58
Loveland	0	15	0	61	56	30	61	62	30
Milford	23	55	17	60	61	25	69	82	38

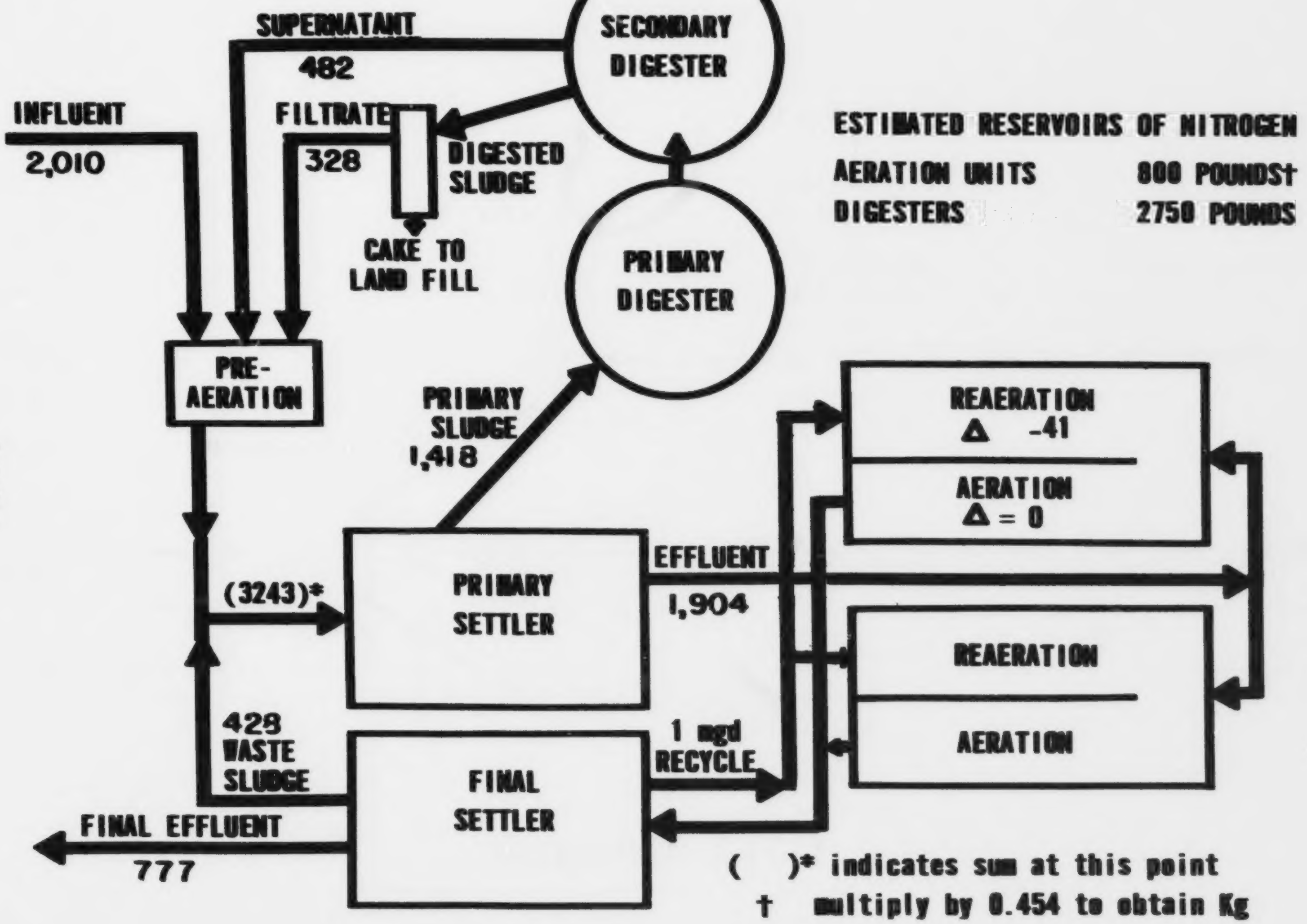
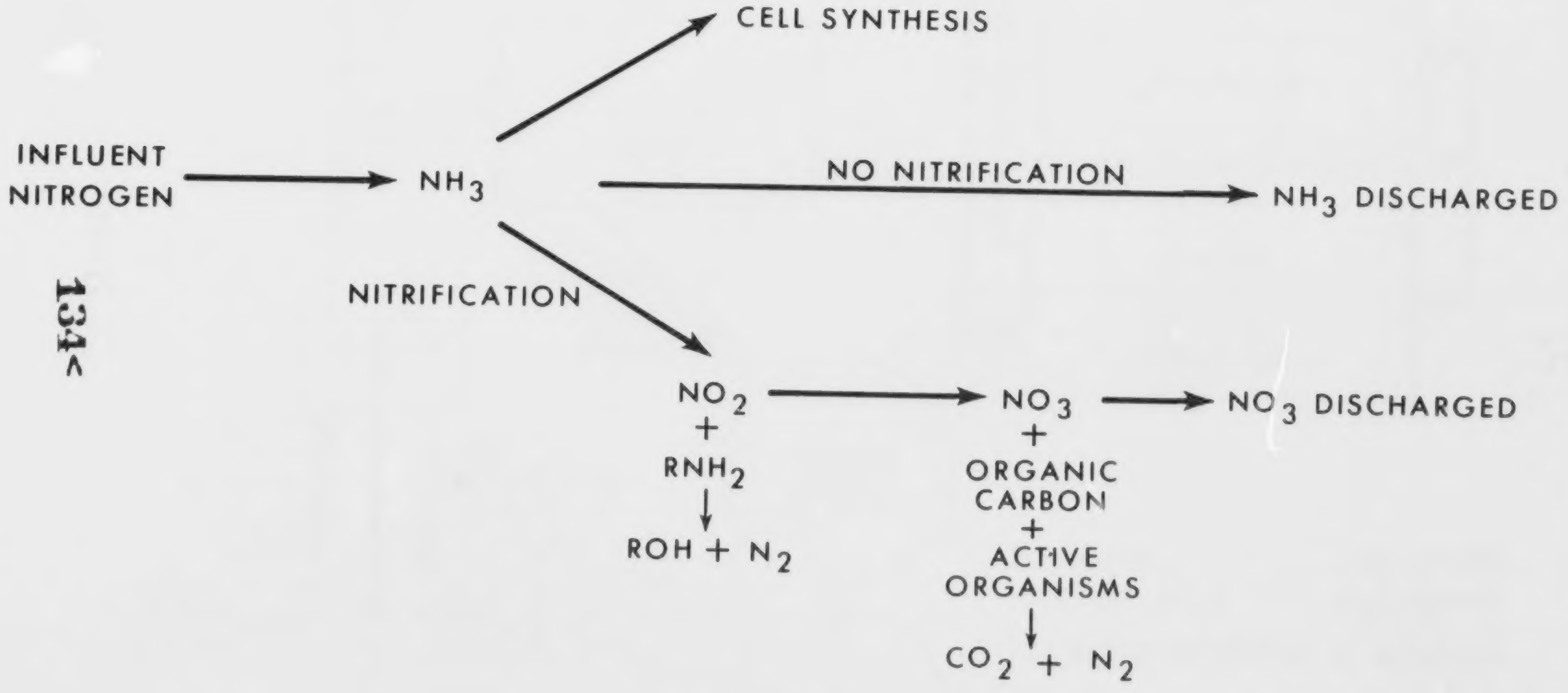
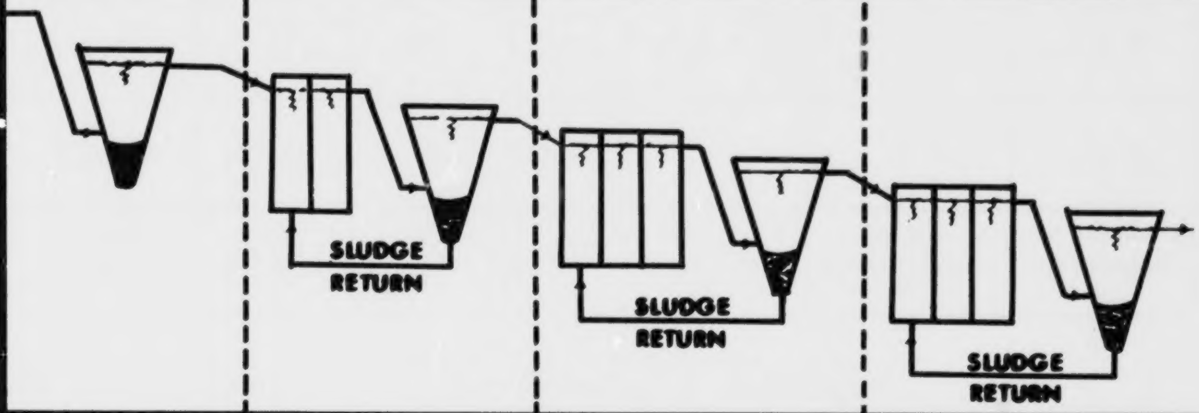

SUPERNATANT
482
INFLUENT
2,010
FILTRATE
328
SECONDARY DIGESTER
DIGESTED SLUDGE
CAKE TO LAND FILL
ESTIMATED RESERVOIRS OF NITROGEN
AERATION UNITS 800 POUNDS†
DIGESTERS 2750 POUNDS
PRIMARY DIGESTER
PRE-AERATION
PRIMARY SLUDGE 1,418
REAERATION Δ -41
AERATION Δ = 0
EFFLUENT
1,904
(3243)*
PRIMARY SETTLER
REAERATION
AERATION
428 WASTE SLUDGE
1 mgd RECYCLE
FINAL SETTLER
FINAL EFFLUENT
777
( )* indicates sum at this point
† multiply by 0.454 to obtain Kg

Figure 1. FIVE-DAY NITROGEN BALANCE: ARCHBOLD, OHIO. ALL VALUES IN POUNDS

NITROGEN CYCLE


INFLUENT NITROGEN
NH3
CELL SYNTHESIS
NO NITRIFICATION
NH3 DISCHARGED
NITRIFICATION
NO2 + RNH2 → ROH + N2
NO3 + ORGANIC CARBON + ACTIVE ORGANISMS → CO2 + N2
NO3 DISCHARGED

PROCESS FLOW DIAGRAM

MAJOR PROCESS FUNCTIONS	REMOVE SETTLEABLE SOLIDS AND PARTICULATE COD	REMOVE BULK OF SOLUBLE ORGANICS AND PRECIPITATE PHOSPHORUS	CONVERT NH_3 TO NO_3	CONVERT NO_3 TO NITROGEN GAS AND CO_2 AND FURTHER REDUCE PHOSPHORUS
CHEMICAL ADDITIVES	NONE	SODIUM ALUMINATE	NONE	SODIUM ALUMINATE AND METHYL ALCOHOL
				
Unit	Primary	High—Rate	Nitrification	Denitrification

SLUDGE RETURN
SLUDGE RETURN
SLUDGE RETURN

MIXED LIQUOR CHARACTERISTICS AFTER ADDITION OF ALUMINUM

	Sludge Age days (Approx.)	MLSS mg/l	MLVSS mg/l	% Volatile	SDI	% Nitrogen V.S. Basis	% Phosphorus V.S. Basis	% Aluminum V.S. Basis
High-Rate Mixed Liquor	2	3,385	1,740	51	1.12	8.3	12.3	11.0
Nitrification Mixed Liquor	22	1,195	770	64	1.15	10.6	8.0	6.6
Denitrification Mixed Liquor	38	2,010	1,385	69	1.46	9.7	8.7	6.0

TABLE IV - Change in Concentration of Various Constituents During Treatment *

Constituent	Raw waste-water mg/l	Primary effluent mg/l	High-rate effluent mg/l	Nitrified effluent mg/l	Denitrified effluent mg/l	Denitrified effluent (filtered) mg/l
COD						
Average	320	218	64	43	44	38
Range	95-1,195	67-566	24-328	17-239	21-194	15-194
Organic-N						
Average	10.3	5.9	0.8	0.4	0.4	0.2
Range	2.3-31	1.4-12	0.8-1.4	0.0-1.0	0.0-1.7	0.0-1.6
Ammonia-N						
Average	11.3	13.7	7.7	0.6	0.3	0.4
Range	0.2-24	2.5-21	0.2-16	0.0-6.0	0.0-2.2	0.0-2.3
Nitrite-N						
Average	--	--	1.1	0.3	0.3	0.4
Range	--	--	0.0-5.8	0.0-2.6	0.0-1.8	0.0-1.9
Nitrate-N						
Average	--	--	4.3	11.5	0.9	0.9
Range	--	--	0.0-10.5	3.7-24.8	0.0-7.5	0.0-6.1
Total phosphorus						
Average	12.6	11.9	2.8	2.6	1.5	1.1
Range	3.0-30	3.0-26	0.4-11.3	0.7-8.2	0.2-5.3	0.2-4.5
S.S.						
Average	157	90	8.8	7.1	7.1	1.7
Range	40-800	28-236	0.0-39	0.0-54	0.0-45	0.0-7
pH						
Mode	7.2	7.3	7.5	7.5	7.7	7.7
Range	6.7-7.6	6.9-7.6	7.2-7.8	7.1-8.0	7.2-8.0	7.0-8.3

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* Each average is result of 40 determinations.

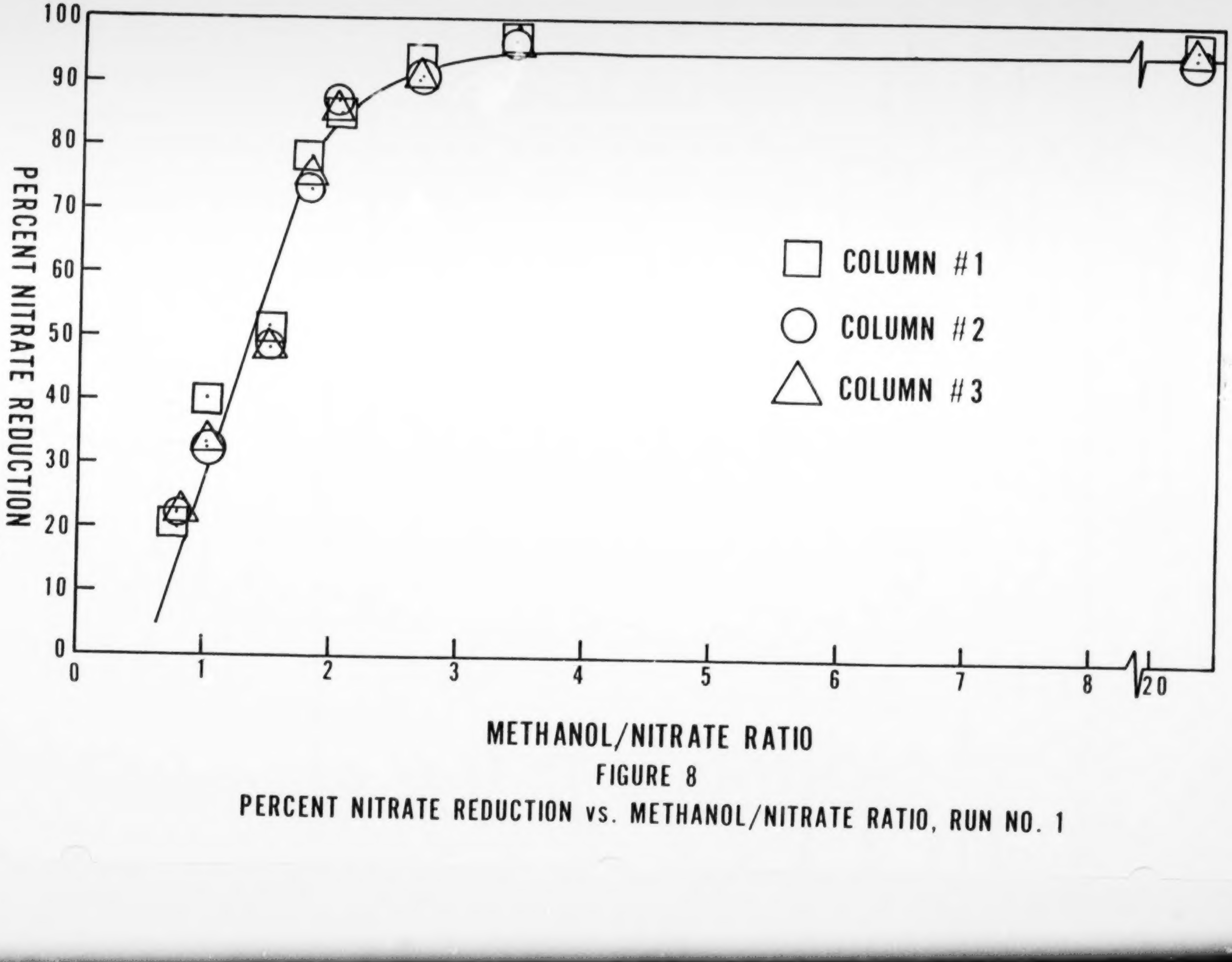


FIGURE 8

PERCENT NITRATE REDUCTION vs. METHANOL/NITRATE RATIO, RUN NO. 1

MODIFICATIONS OF THE DENITRIFICATION PROCESS

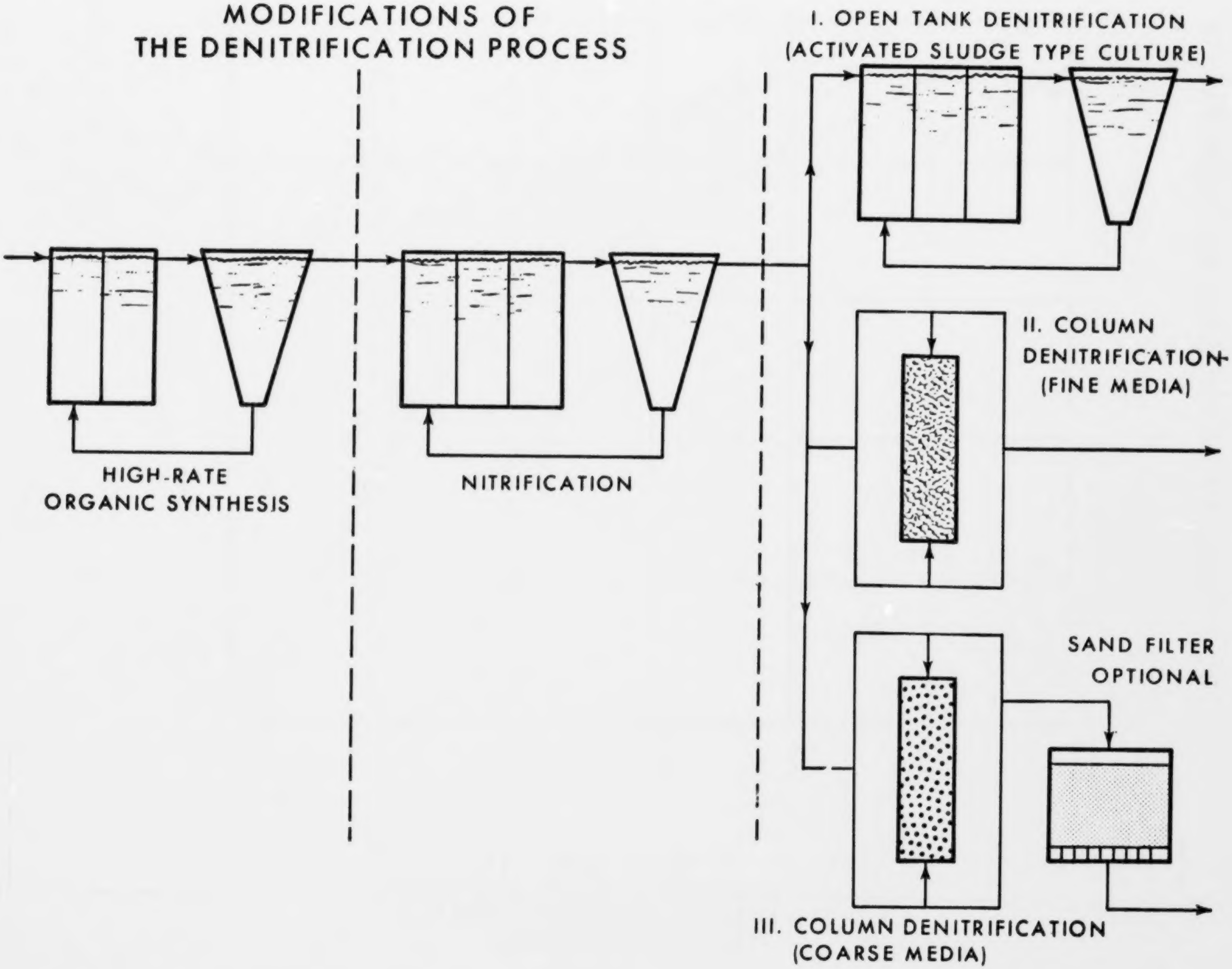
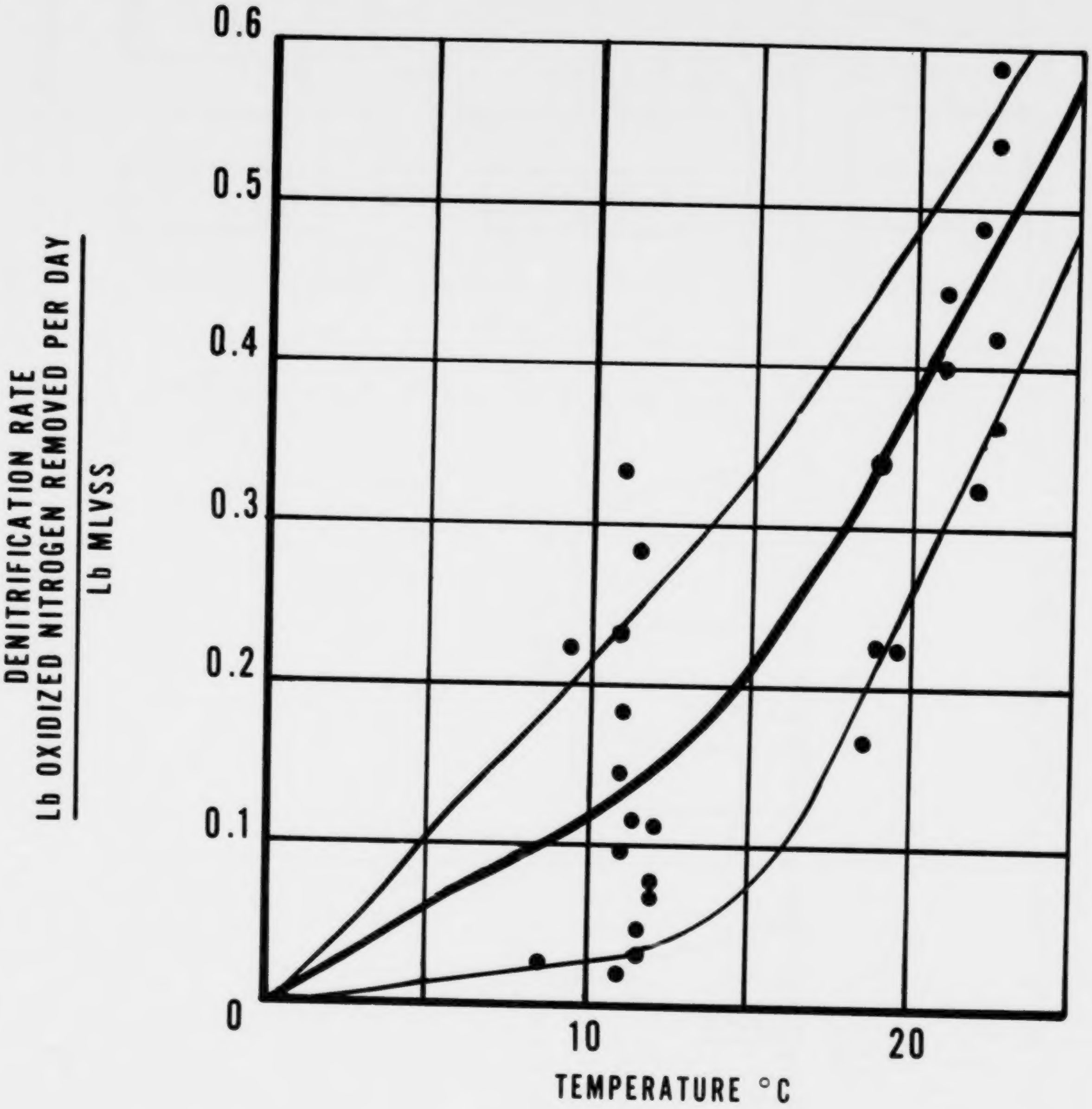

I. OPEN TANK DENITRIFICATION
(ACTIVATED SLUDGE TYPE CULTURE)
II. COLUMN DENITRIFICATION
(FINE MEDIA)
SAND FILTER OPTIONAL
HIGH-RATE ORGANIC SYNTHESIS
NITRIFICATION
III. COLUMN DENITRIFICATION
(COARSE MEDIA)

FIGURE 7 - DENITRIFICATION RATE VS. TEMPERATURE

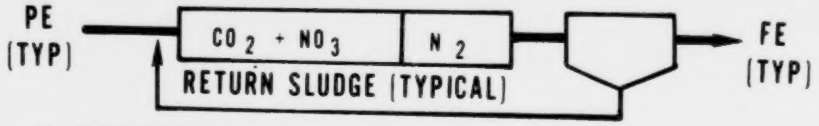


Reference: M. Mulbarger

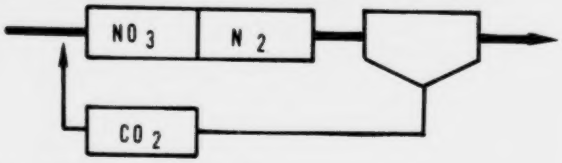
FIGURE 1
PROCESS ALTERNATIVES FOR NITRIFICATION - DENITRIFICATION

I. SINGLE SLUDGE SYSTEMS

A. EXTENDED AERATION

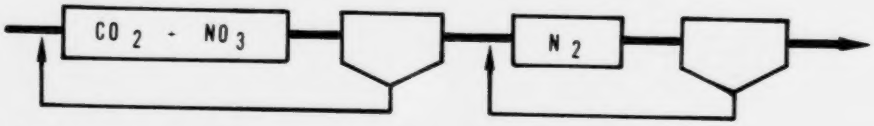

PE
(TYP)
CO 2 + NO 3
N 2
FE
(TYP)
RETURN SLUDGE (TYPICAL)

B. CONTACT STABILIZATION MODIFICATION

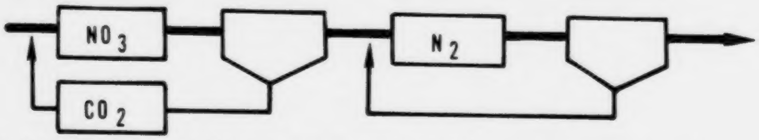


II. DUAL SLUDGE SYSTEMS

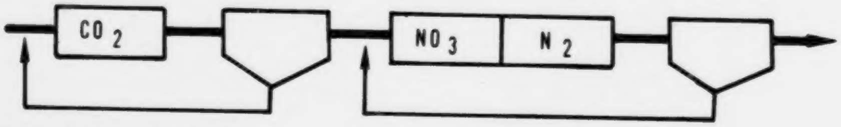
A. EXTENDED AERATION WITH ISOLATED DENITRIFICATION



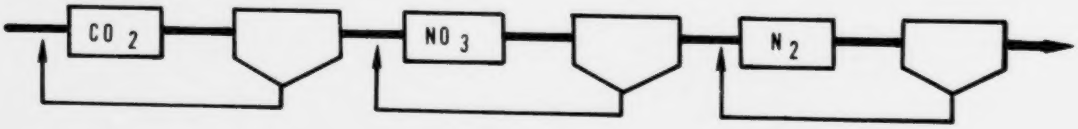
B. CONTACT STABILIZATION MODIFICATION WITH ISOLATED DENITRIFICATION



C. HIGH RATE WITH COMBINED NITRIFICATION - DENITRIFICATION



III. THREE SLUDGE SYSTEM



CODE

CO_2 CARBON REMOVAL (OXIDATION + SYNTHESIS)

NO_3 NITRIFICATION

N_2 - DENITRIFICATION

Reference: M. Mulbarger

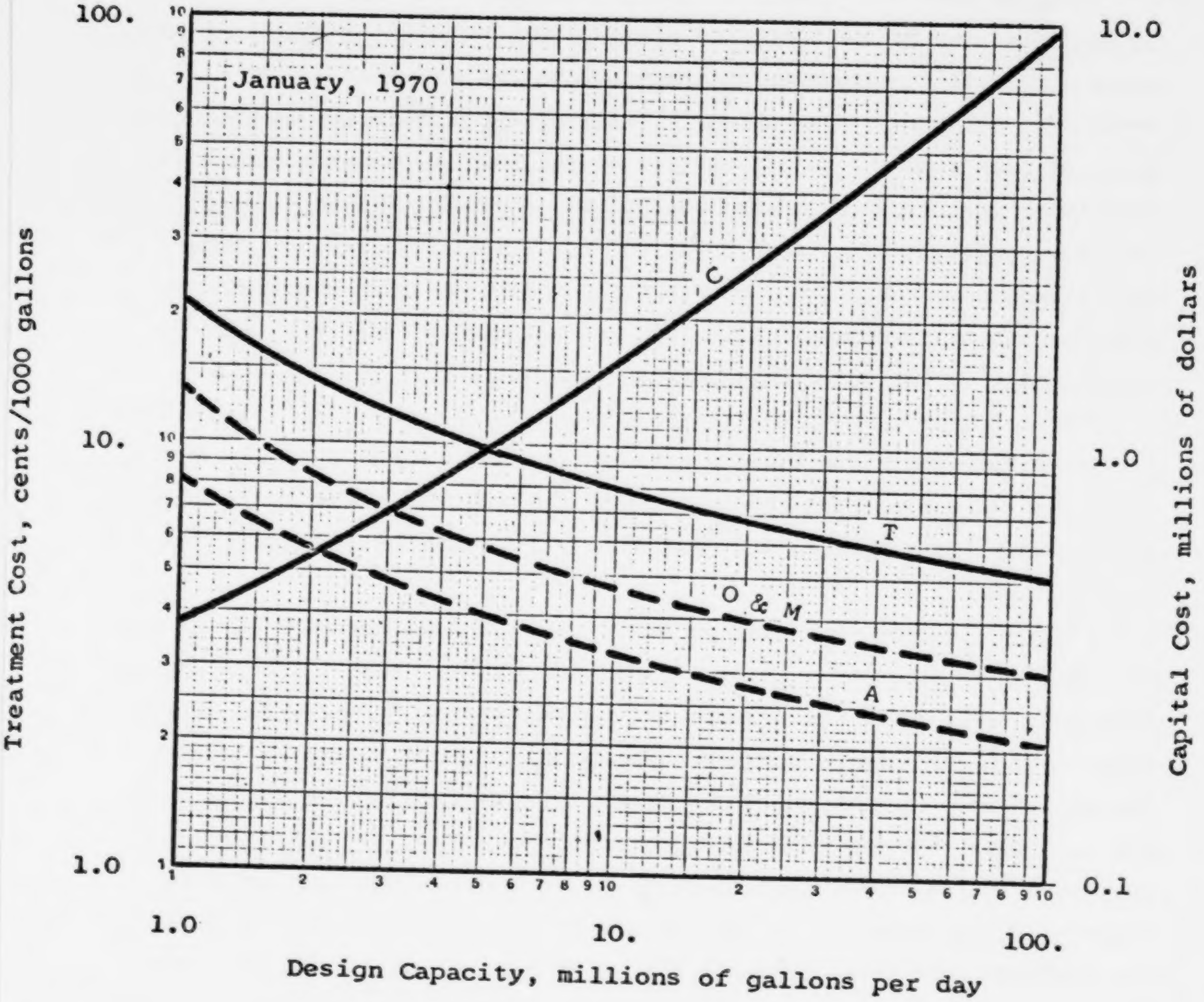
Capital Cost - 10 mgd Plant - Millions of Dollars
(1969-1970 Dollars)

System Item	Standard Activated Sludge	IA	IB	IIA	IIB	IIC	III
Pre-Treatment Control Bldg. primary and final clarification, and disinfection.	1.5	1.5	1.5	1.5	1.5	1.5	1.5
Sludge Thickening, dewatering, treatment and disposal.	2.6	2.2	2.2	2.2	2.2	3.0	3.0
Aeration	1.1 1.4 5.2 5.5					0.3 0.5 4.8 5.0	0.3 0.5 4.8 5.0
Nitrification		1.9 2.7	1.1 1.5	1.9 2.7 5.6 6.4	1.1 1.5 4.8 5.2	0.9 1.1	0.9 1.1 5.7 6.1
Denitrification		0.5 0.7 6.1 7.1	0.5 0.7 5.3 5.9	0.9 1.1 6.5 7.5	0.9 1.1 5.7 6.3	0.5 0.7 6.2 6.8	0.9 1.1 6.6 7.2

Reference: M. Mulbarger

NITRIFICATION AND DENITRIFICATION IN DISPERSED GROWTH REACTORS

Capital Cost, Operating & Maintenance Cost, Debt Service
versus
Design Capacity



C = Capital Cost, millions of dollars
A = Debt Service, cents per 1000 gallons (6% and 25 yr.)
O & M = Operating and Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

Robert Smith
4676 Columbia Parkway
Cincinnati, Ohio 45226

Paper No. 3

NITROGEN REMOVAL FROM WASTEWATERS BY COLUMN REACTORS

J. M. Smith

Once nitrates are formed in sewage treatment they can be converted to nitrogen gas by the action of microbes and a source of BOD. Columnar nitrate reduction represents an alternate to the suspended growth systems as a means of biochemically reducing the nitrate ion to elemental nitrogen (see Figure 3). A variety of facultative bacteria common in wastewater including the genera Pseudomonas, Achromobacter and Bacillus are known denitrifiers. When supplied with a suitable organic carbon source under anaerobic or near anaerobic conditions these organisms use nitrates and nitrites as the terminal hydrogen acceptors during the fermentation of the added substrate to ultimately form nitrogen gas.

In a packed column the cell residence time of the surface bound slime is much greater than the hydraulic detention time. This, combined with a large contact surface and short diffusion distances afforded by small media such as sand, provides an efficient system for rapid denitrification of an applied feed.

Work at Lebanon, Ohio (J. M. Smith, 1970, Unpublished) has shown that the smaller media systems (sand to 3/4 inch diameter stone) are effective when operated downflow at surface loading rates of 7.0 gpm/ft^2 and at actual contact times of 5 to 30 minutes. The downflow operation, using either granular carbon or sand, requires daily backwashing using 1-3 percent of the throughput stream to relieve pressure drop due to the accumulation of suspended solids in the upper portion of the column. The denitrifying slime is firmly attached to the media surface, and is not removed during the backwash operation. Greater than 90 percent nitrate reduction can be achieved within these columns at contact times of 10 minutes for sand and 30 minutes for the 3/4 inch stone. The effluent normally contains less than 2.0 mg/l of nitrate nitrogen with effluent turbidities less than 3 JTU, indicating little solids contribution from the attached organisms. The ammonia and organic nitrogen content of the feed is not significantly changed during the denitrification process.

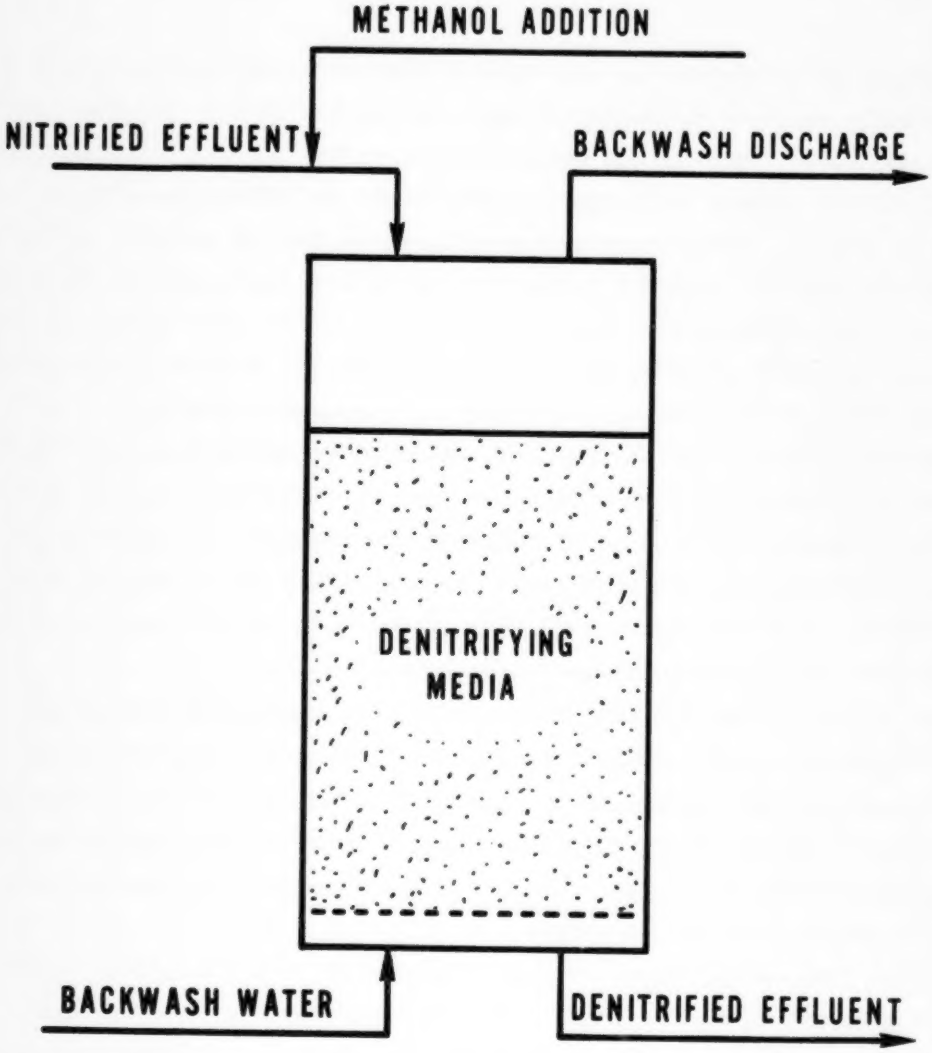

METHANOL ADDITION
NITRIFIED EFFLUENT
BACKWASH DISCHARGE
DENITRIFYING MEDIA
BACKWASH WATER
DENITRIFIED EFFLUENT

FIGURE 3

PACKED BED DENITRIFICATION REACTOR

Larger media varying in size from 1 inch to 2 inch aggregate has been successfully employed to denitrify agriculture subsurface drainage (Tamblyn and Sword, 1969). The larger media permits upflow operation without backwashing at the expense of longer contact times and increased effluent suspended solids. Nitrate reduction of greater than 90 percent can be achieved in contact times of 1 hour for the 1 inch aggregate and 2 hours for the 2 inch aggregate at temperatures above 12° C. The 2 inch columns have been operated continuously for over six months on agriculture subsurface drainage without the loss of efficiency or solids accumulation.

Methanol is used as the supplemental organic carbon source of choice for denitrification because of its low cost, biodegradability and ease of handling. Approximately 3 mg of methanol are required per mg of NO_3-N removed including the requirement for deoxygenating the nitrified feed. The chemical cost for removing 20 mg/l of NO_3-N in the presence of 5 mg/l of dissolved oxygen is estimated to be 1.6¢/1000 gal.

While the packed bed provides an efficient contacting system for denitrification, total nitrogen removal via biological nitrification-denitrification will depend on the combined efficiency of the nitrification stage as well as the denitrification stage. This interdependence of two biological systems will limit total nitrogen removal efficiencies to a practical value of about 90 percent.

Pilot plant experiences to date firmly support a full scale demonstration of columnar nitrate reduction at the 1 MGD level or above. One possibility is the construction of a new facility that incorporates a two stage nitrifying system followed by columnar denitrification. An alternate, and equally plausible idea is to modify an existing conventional activated sludge process by splitting the aeration tanks into a two sludge system by the addition of a second settler, perhaps a tube settler, followed by denitrification on a relatively fine media that will also serve as a solids polishing process. The design of a denitrification system is straightforward and involves basic principles of biological oxidation and filtration. When furnished with basic denitrification data any competent sanitary engineering firm should be able to successfully design a treatment plant capable of removing nitrogen.

St. Armant and McCarty (1969) have demonstrated the use of anaerobic filters for removal of nitrates from irrigation return flow collected in drain tiles. They found that methanol requirements were close to the theoretical. The most satisfactory media was one inch diameter gravel operated upflow. Detention times of one hour were satisfactory at temperatures above 12°C. The treated water contained less than 1 mg/l nitrate and slightly more nitrite. It had no dissolved oxygen. Sludge production by the denitrifying organisms produced turbidity which would have to be removed by filtration.

Paper No. 4

AMMONIA NITROGEN REMOVAL BY STRIPPING WITH AIR

J. B. Farrell

Ammonia is a gas which dissolves in water to an extent controlled by the partial pressure of the ammonia above the water. An increase in its partial pressure causes more ammonia to dissolve, reducing its partial pressure causes ammonia to leave the water phase and enter the gas phase. If it is desired to remove ammonia from water, this can be done by contacting the water with a sufficient quantity of ammonia-free air. This physical process is called desorption or, more commonly, "stripping."

Wastewater contains the ammonium ion at concentrations which typically range from 15 to 25 mg/l. If the ammonium ion is converted to dissolved ammonia by raising the pH, the ammonia may be stripped out by contacting the wastewater with air. Early studies of feasibility of stripping of ammonia from wastewater (Kuhn, 1956) showed a major difficulty: the volume of air per unit volume of water is very high, about 400 cubic feet per gallon of water in a countercurrent-flow packed tower. The pressure drop across a packed tower is about 1.0 psi or 28" of water. Since power is proportional to the product of volumetric flow rate and pressure drop, the cost for power made the process prohibitive even at this relatively low pressure drop.

The problem of high power cost was solved by investigators at the South Lake Tahoe Public Utility District (Slechta and Culp, 1967) who used a slat-filled tower such as is used for cooling water to contact water and air. The pressure drop across such a device is very low--about 1/2 inch of water--so power costs are reduced to reasonable levels. Removal efficiencies as high as 90 percent were obtained in a 24-foot high tower in which water and gas were contacted in a nearly countercurrent fashion. On the basis of this experience, a full scale stripping tower was constructed at South Lake Tahoe. The tower was designed to remove 90 percent of the ammonia from 3-1/2 MGD of Tahoe's renovated wastewater. The air flow is not countercurrent to the water but flows across the tower (cross-flow), while the water drips downward through the packing.

Initial operation of Tahoe's stripping tower was in the winter and immediately revealed a limitation of ammonia stripping. The cold air contacting the wastewater quickly cooled the water. Ammonia solubility is higher in cold water than in warmer water; consequently, more air is required to remove it. For example, at 0° C it would take about 800 cubic feet of air per gallon of water to remove 90 percent of the ammonia from wastewater in a countercurrent tower. Since the Tahoe tower was designed for 400 cubic feet per gallon, percent removal was much lower than 90 percent. When air temperature fell below 0°, freezing of water occurred at the air inlets, making the tower inoperable. From theoretical considerations, difficulty with freezing may be anticipated when the wet-bulb temperature of the ambient air is below 0° C.

Another problem which developed at Tahoe is the formation of scale in the tower. The scale is chiefly calcium carbonate. It forms because the previously lime-treated effluent is supersaturated with respect to calcium carbonate. A more important reason is the presence of carbon dioxide in the air. A small portion of the carbon dioxide reacts with the basic wastewater, forming calcium carbonate which precipitates. In the case of the tower at Tahoe, the sludge can be flushed from the tower except from inaccessible areas which cannot be reached with a water jet. A pilot plant ammonia stripping tower at Blue Plains, Washington, D. C., had similar scaling problems, except the scale is hard and adheres to the tower fill. They have been able to get a soft scale by reducing the pH to 10. A pilot tower in New England had similar hard scale (Sawyer, 1970). However, a pilot plant tower has been run in South Africa (van Vuuren, 1970), and here the scale has been so soft that the force of the water dripping through the tower fill has been sufficient to flush out the scale.

The causes of the differences in the nature and amount of scale in various locations has not been resolved. Some suggested factors influencing the nature of the scale are: type of contact (countercurrent or cross-flow), recirculation of sludge, pH of the wastewater, and chemical makeup of the wastewater. Studies are in progress to see if the scale can be prevented from forming, or if it can be made nonadherent.

The cost of ammonia stripping has been estimated by Suhr (1969) for South Lake Tahoe facility to be about 2.9¢ per 1000 gallons of wastewater treated. This does not include the cost of the lime and facilities to raise the pH to about 11. These costs have been charged to phosphorus removal because this is the direct objective of the lime addition. If 90% removal of ammonia nitrogen is required even in cold weather, these costs should be increased by about 50% to provide for a higher air-to-water ratio. In below-freezing weather the tower will be inoperable. It is possible to heat the incoming air to temperatures above 0° C and add moisture to it to prevent freezing. This procedure would only be economical if subfreezing days are rarely encountered.

Methods for designing ammonia stripping towers have been developed at the Taft Center by Roesler et al. (1970) but have not yet appeared in the literature. Ammonia stripping is feasible when the temperature is above freezing but there is danger of serious fouling by scale. The best approach for minimizing scale and its effects appears to be to use a pH of about 10.5, countercurrent operation rather than cross-flow, and an open fill to allow for easy flushing of accumulated solids.

In order to make more information available than is presented in the above summary, reproductions of some of the slides used in the oral presentation given at Dallas, Texas have been attached.

VAPOR PRESSURE OF A GAS DISSOLVED IN WATER

Henry's Law

At low concentrations, Henry's Law holds:

$$\text{Vapor Pressure of Dissolved Substance} = H \cdot \text{Concentration of Dissolved Substance}$$

Alternative Forms:

$p_a = H \cdot c_a$ where p_a = vapor pressure

c_a = concentration

H depends on temperature

$y = k \cdot x$ where y and x are mol fractions of A in vapor and liquid

k depends on temperature and atmospheric pressure

COMPARISON OF SOLUBILITY OF CARBON DIOXIDE AND NH_3 IN WATER

TEMPERATURE	HENRY'S LAW CONSTANT (H)	
(°C)	CO_2	NH_3
20	1,420	0.7
40	2,330	1.8
60	3,410	3.8

WHERE $p_a = Hx_a$

p_a = VAPOR PRESSURE OF DISSOLVED SUBSTANCE "a" (atm)

x = MOLE FRACTION OF "a"

H = HENRY'S LAW CONSTANT

MINIMUM AIR REQUIREMENT FOR STRIPPING NH_3 AND CO_2 FROM WATER AT 20°C IN A COUNTERCURRENT OPERATION

Stripping NH_3 out of basic solution*		Stripping CO_2 out of acidic solution*	
NH_3 in water	NH_3 in air	CO_2 in Water	CO_2 in air
20 mg/l ↓	↑ 8.7×10^{-6} g/g	20 mg/l ↓	↑ 17.6×10^{-3} g/g
2 mg/l	0	2 mg/l	0
Air Required: 225 cu ft/gal		0.11 cu ft/gal	

*Entering water stream and exiting air stream are assumed to be in equilibrium

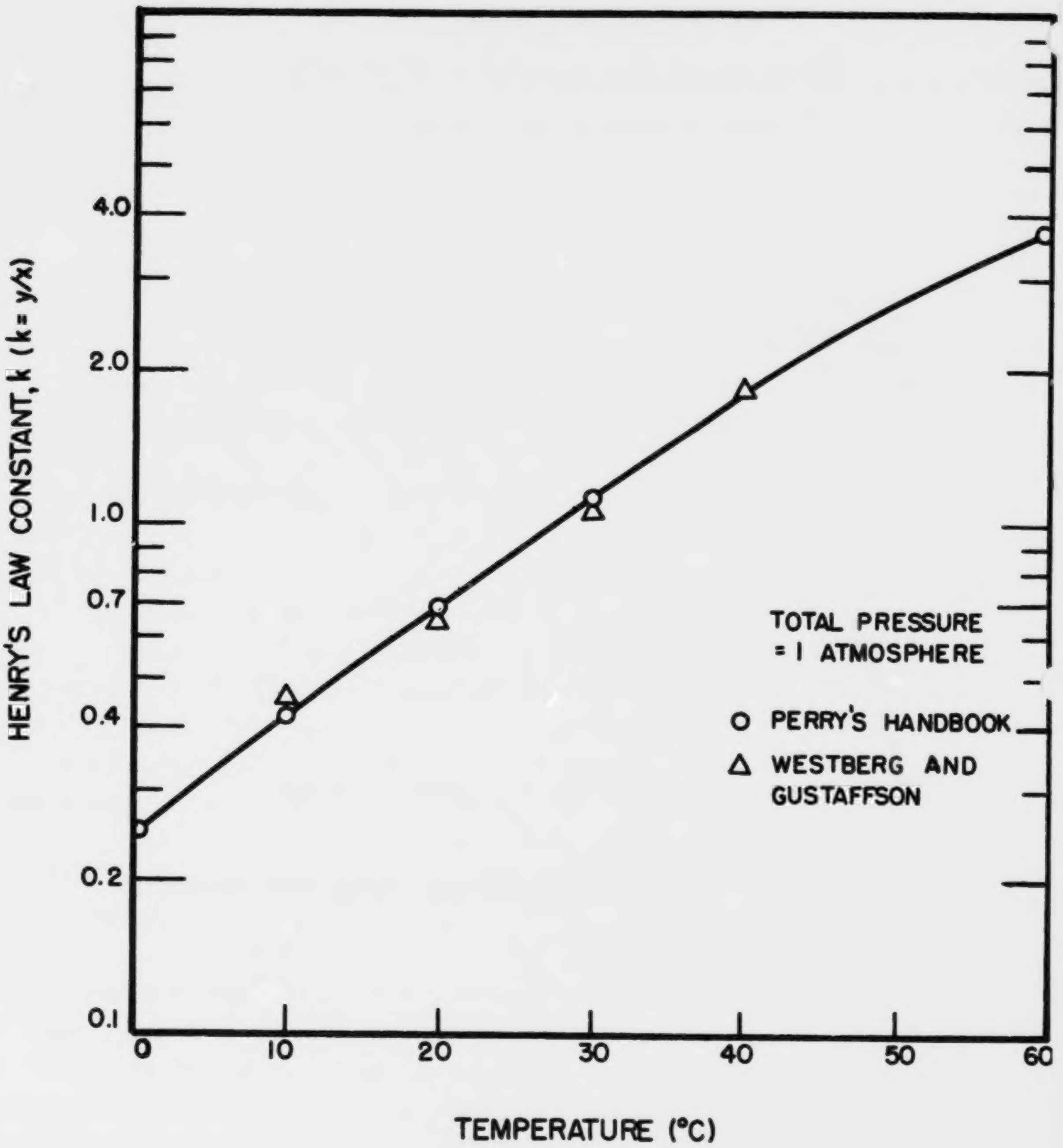


Figure 4. Effect of Temperature on Henry's Law Constar for Ammonia–Air–Water System.

Paper No. 5

REMOVAL OF AMMONIA NITROGEN BY SELECTIVE ION EXCHANGE

R. B. Dean

Ammonia exists in solution predominantly as the ammonium cation (NH_4^+) unless the pH is higher than about 9.5. The ammonium ion is very similar to the potassium ion in size and is precipitated by the same reagents that precipitate potassium. Adsorption of ammonium ion on ion exchange resins is ordinarily very similar to the adsorption of potassium or sodium ions. Therefore, conventional water softening ion exchange resins which are selective for calcium and magnesium do a relatively poor job of removing ammonium from dilute solutions. Total deionization by mixed bed ion exchange resins will, of course, remove ammonium ions along with other cations but this process is too costly for wastewater treatment.

Certain zeolites show unusual selectivity for the ammonium ion. A number of these have been investigated by the Atomic Energy Commission because they also show selectivity for cesium and potassium ions. A demonstration project at the Battelle Memorial Institute - Pacific Northwest (Hanford) Laboratories, 1969 (see Mercer, 1970), showed that certain zeolites, including the naturally occuring mineral clinoptilolite, had a high selectivity for ammonium in natural and wastewaters. A trailer mounted demonstration plant with a capacity of 100,000 gallons per day was built as a cooperative demonstration project between the FWQA and Battelle-Northwest. This trailer is now operated under contract to the FWQA to demonstrate selective ion exchange removal of ammonium ions from solution.

Preliminary work demonstrated that native clinoptilolite from Hector, California, could be used after grinding and sieving to 20 x 50 mesh. When columns filled with this material were fed secondary effluent they clogged rapidly; therefore, the trailer incorporates an inclined-tube settler clarifier to remove turbidity from the feed. Clarified water is treated by conventional downflow through two columns in series. When ammonia breaks through the second column, the first column is taken offstream and a third column containing freshly regenerated clinoptilolite is added after the second column which now becomes the first in the series. The loaded column is regenerated by backwashing with limewater containing sodium chloride to speed up the rate of regeneration. The concentration of sodium chloride is approximately that which will be reached after several regeneration cycles. The high pH of the limewater converts the ammonium ion to unionized ammonia gas in solution. This is allowed to build up to a level of about 500 mg/l by recycling of the limewater. Precipitates such as magnesium hydroxide which may form in some hard waters are removed by settling.

The ammonia laden limewater is then pumped through a packed column through which heated air is blown to remove the ammonia. The problems of ammonia dispersal to the atmosphere are similar to those encountered in direct air stripping of ammonia. However, the ammonia content of the gas stream is 20 to 50 times as strong as occurs with direct air stripping and there is proportionately much less air to be heated in cold weather. On a large scale it would probably be more economical to use stripping towers of the cooling tower type rather than packed towers because the latter have very high pressure drops. However, a packed tower is much more compact and could be incorporated in the available space on the trailer.

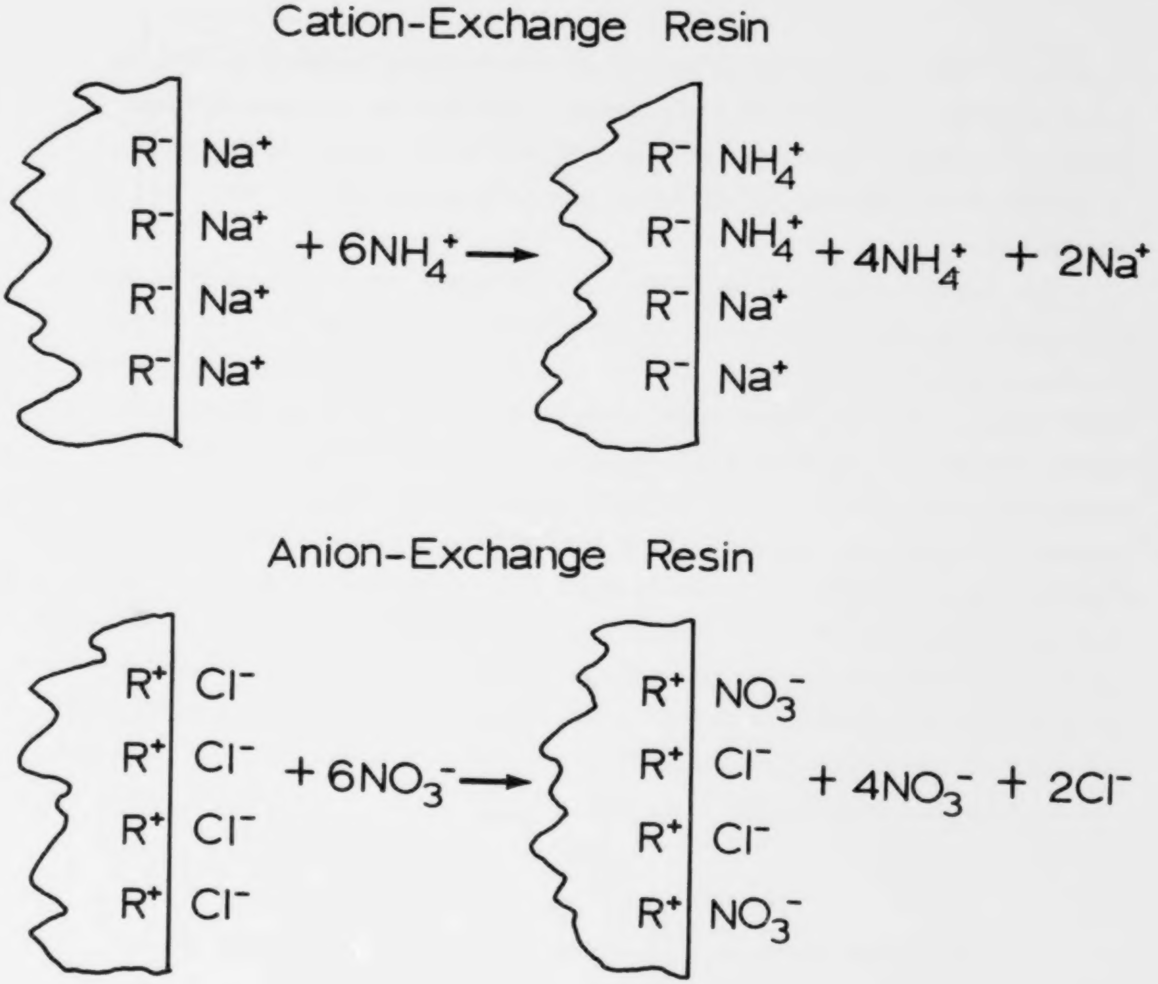
Because of the high concentration of ammonia, its odor is readily detectable in the effluent gases. Of course, the total amount of ammonia vented to the atmosphere is the same whether ammonia is stripped directly or is first concentrated in an ion exchange column. Direct stripping of ammonia from municipal effluents would produce a concentration of ammonia nitrogen in the air of approximately 0.03 ppm, which is well below the threshold for detection by odor.

Ammonia in the atmosphere reacts with sulfur dioxide which is present in large excess over all cities. If all of the ammonia present in sewage in the United States were to be discharged to the atmosphere, it would be equivalent to less than 10 percent of the sulfur dioxide being discharged to the atmosphere. Ammonia and SO_2 together can cause fog under some conditions; however, unneutralized sulfur dioxide eventually forms sulfuric acid which is even more objectionable. Further studies on the behavior of ammonia in urban atmospheres are warranted (Miner, 1969).

Ammonia gas in the atmosphere, whether or not it reacts with sulfur oxides, is adsorbed by moisture in soil, vegetation, and open water surfaces. That portion which goes into the soil and vegetation will be bound and will no longer be a pollutant. However, if prevailing winds would blow ammonia laden air over inland bodies of water, as for example at Toledo, then ammonia stripping to protect surface waters would not be appropriate.

The cost of removing ammonia by ion exchange has been estimated by Eliassen as 8.1¢/1000 gallons. His estimate includes the cost of brine for regeneration but does not include the cost of disposal of the ammonia laden brine. In most areas this could be a serious problem since discharge to surface waters would be self-defeating. Pilot studies at Battelle-Northwest indicated a cost of between 3 and 6¢/1000 gallons for their Zeolite method. An estimate prepared for a 7.5 MGD plant at Lake Tahoe showed 15¢/1000 gallons which is probably high. By comparison with other fixed bed adsorption methods it is reasonable to suggest a cost of about 10¢/1000 gallons for ammonia removal by zeolites. More reliable cost estimates will be available at the conclusion of present pilot plant studies. Alternative methods of regenerating zeolites which convert the ammonia to nitrogen gas are also being investigated.

The above paper was delivered at Dallas, Texas by J. B. Farrell. In order to make more information available than is presented in the paper, a number of the slides used in the oral presentation have been attached below.



Cation-Exchange Resin

Anion-Exchange Resin

R⁻ Na⁺
R⁻ Na⁺
R⁻ Na⁺
R⁻ Na⁺
+ 6NH₄⁺ →
R⁻ NH₄⁺
R⁻ NH₄⁺
R⁻ Na⁺
R⁻ Na⁺
+ 4NH₄⁺ + 2Na⁺
R⁺ Cl⁻
R⁺ Cl⁻
R⁺ Cl⁻
R⁺ Cl⁻
+ 6NO₃⁻ →
R⁺ NO₃⁻
R⁺ Cl⁻
R⁺ Cl⁻
R⁺ NO₃⁻
+ 4NO₃⁻ + 2Cl⁻

FIGURE 1

REMOVAL OF IONS FROM AQUEOUS SOLUTION
BY MEANS OF ION-EXCHANGE RESINS

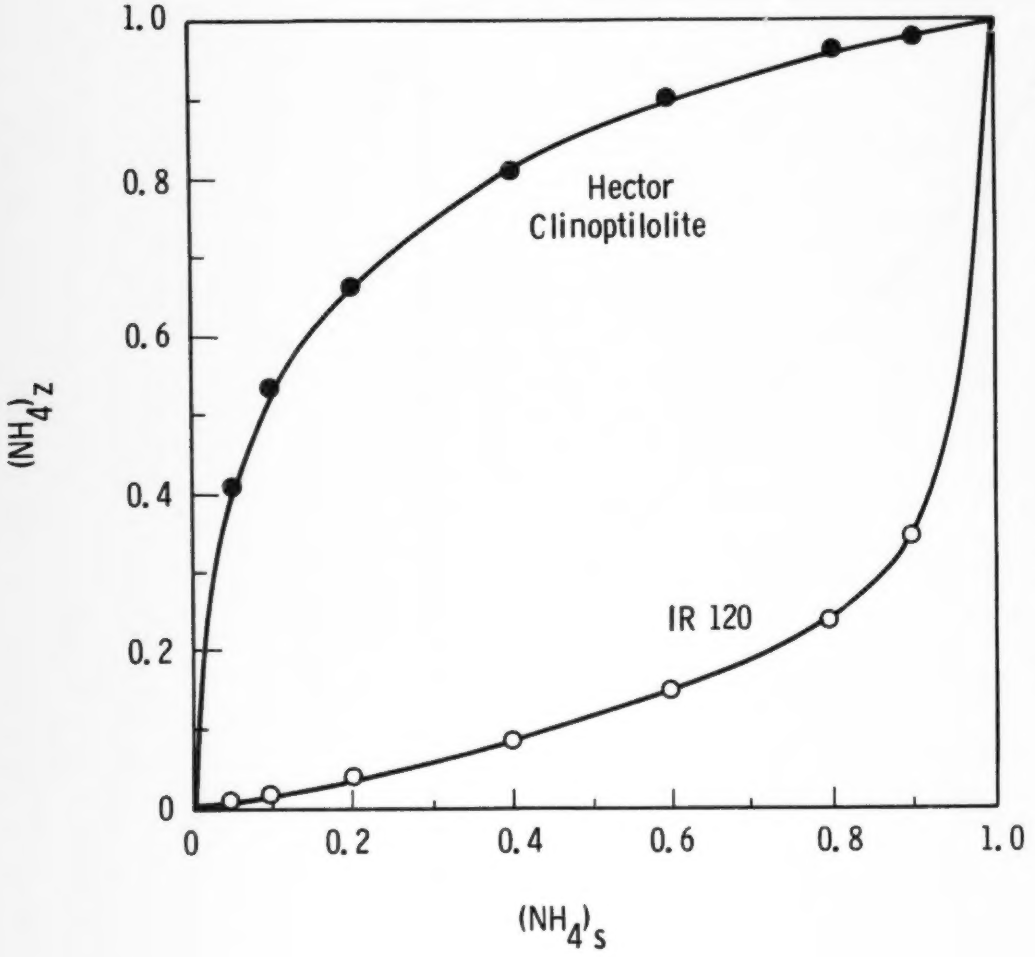


Figure 2. The 23°C isotherms for the reaction, $(Ca)_z + 2(NH_{4_N}) = 2(NH_4)_z + (Ca_N)$ with Hector clinoptilolite and IR 120. Total equilibrium solution normality was constant at 0.1N$(Ca)_z$, $(NH_4)_z$ = equivalent fraction of calcium or ammonium on the zeolite. (Ca_N), (NH_{4_N}) = normality of calcium or ammonium in the equilibrium solution. The ammonium capacity of IR 120 was 4.29 meq/g of air-dried resin.

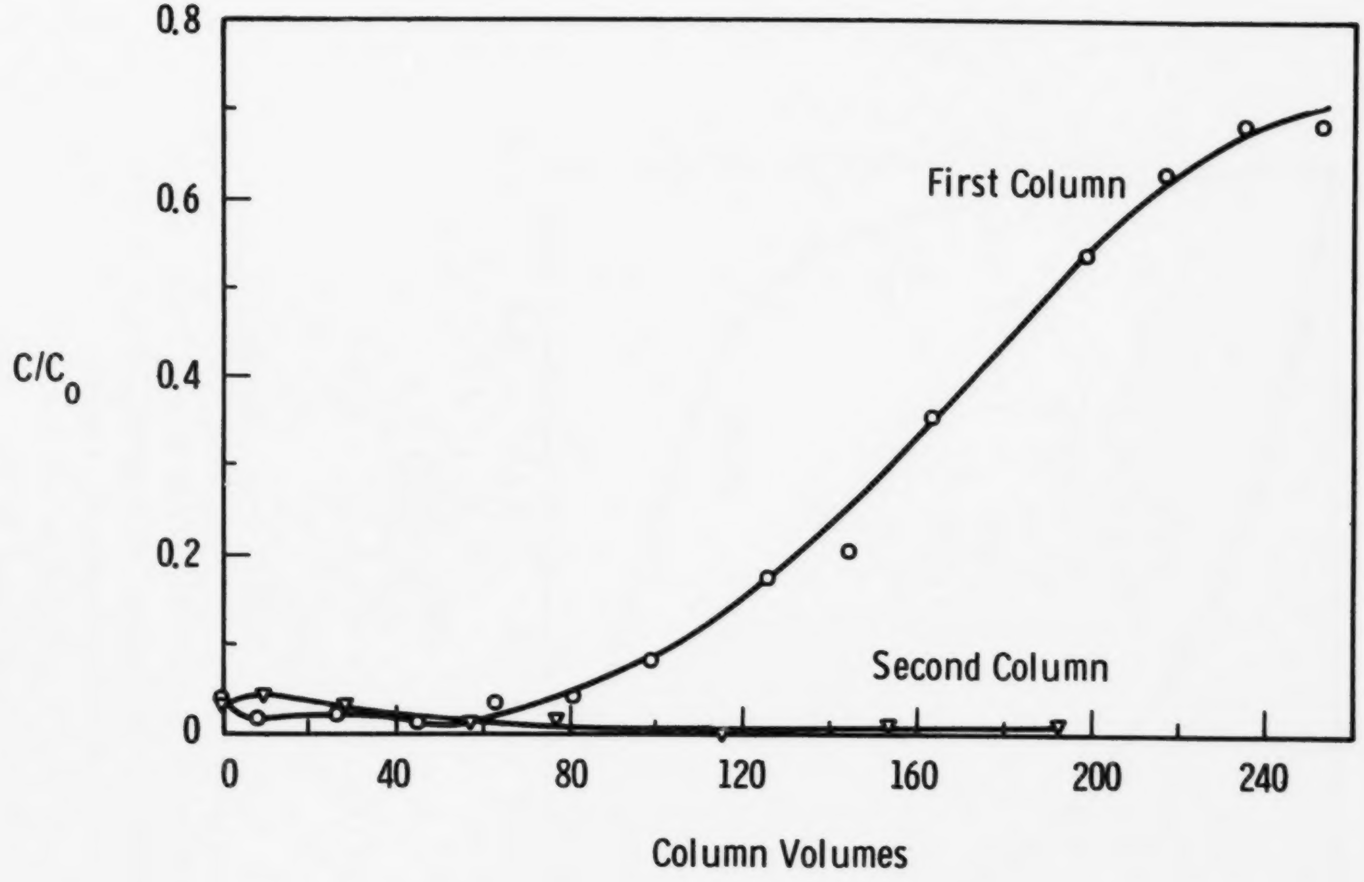


Figure 11. Ammonia Breakthrough for Two Clinoptilolite Columns in Series

Flow rate	20 cv/hr.
Column dimensions	2" x 24".
Clinoptilolite grain size	20 x 50 mesh.
Temperature	26°C.
Feed	Alum coagulated and filtered Richland Effluent NH_3 - N content 18.5 mg/l.
Previous regeneration	0.015N NaOH at 10 cv/hr, 40 cv total.

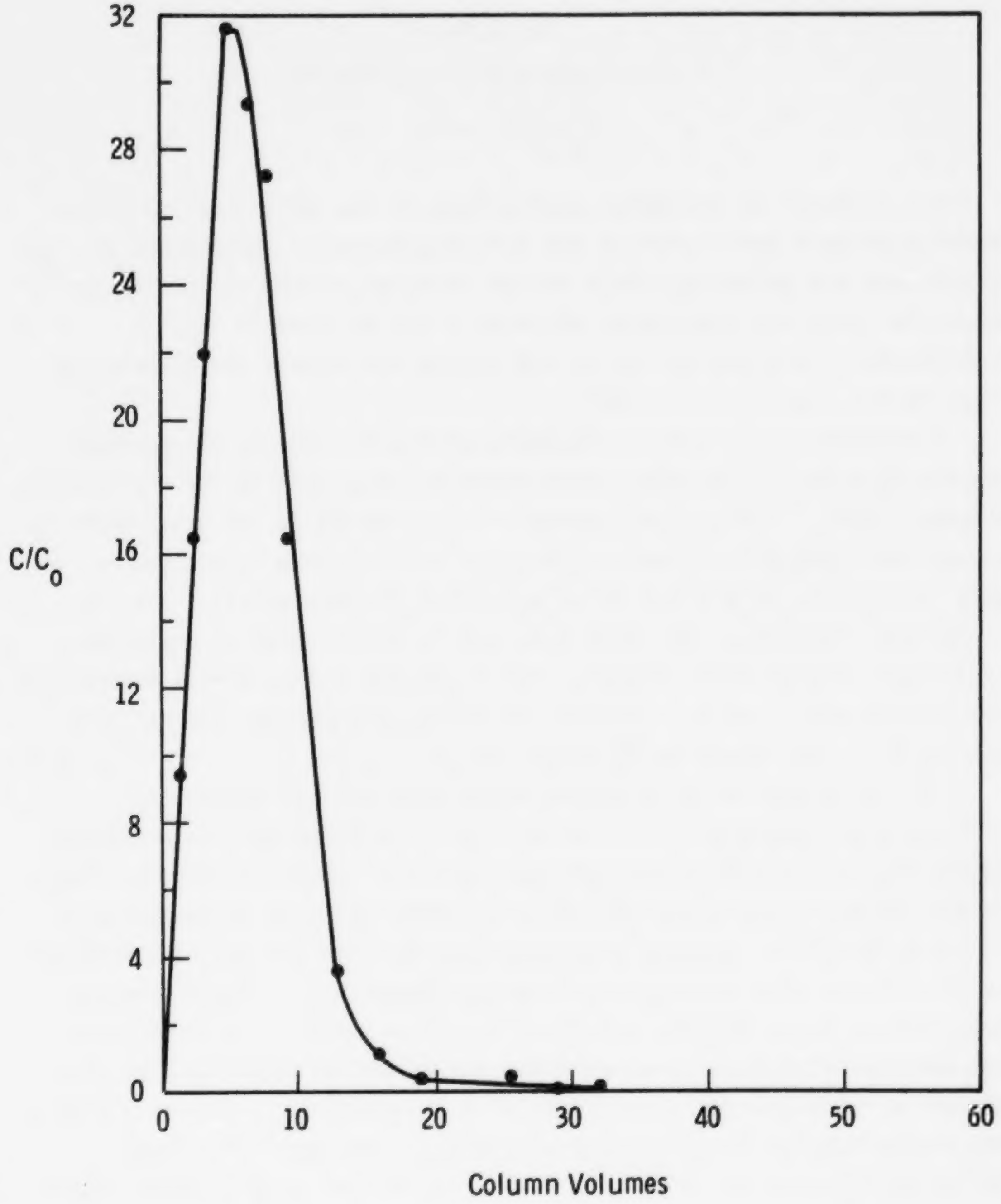


Figure 12. Ammonia Elution with Recycled Regenerant Saturated with Lime.

Flow rate . . . . 10 cv/hr upflow.
Column dimensions . . . . 2" x 24".
Clinoptilolite. . . . . 20 x 50 mesh.
Temperature . . . . 26°C.

Paper No. 6

ION EXCHANGE FOR NITRATE REMOVAL

R. A. Dobbs

Ion exchange is a standard unit process in the water treatment field. Large units have been installed for softening municipal water supplies. The resins used are fairly selective for the divalent cations of calcium and magnesium which are responsible for most of the hardness in water. Unfortunately, a comparable resin with a high selectivity for nitrate ion over other anions does not exist.

A technical and economic evaluation of the feasibility of removing nitrate from irrigation return drain water has been made by the Dow Chemical Company (1968). Although the concentration of anions in the drain water is considerably higher than the corresponding concentration in wastewater, the ratio of nitrate to chloride is quite similar for both waters. Several conclusions related to the drain water can be extrapolated to municipal wastewater. Commercially available anion exchange resins have a selectivity for nitrate over chloride. In order to develop an economic process this selectivity ratio should be approximately 20.

Even if resins of the required selectivity could be developed, additional problems must be solved in order to make the process practical. Regeneration of selective ion exchange resins are usually difficult. Based on the nitrate anion alone, efficiency of sodium chloride regeneration is of the order of 1%. Another serious problem involves the ultimate disposal of the nitrate laden brine produced during regeneration. Theoretically, resin regeneration could be achieved with sodium sulfate. Nitrate could be reduced chemically with ferrous sulfate and neutralized with lime to produce sludges and nitrogen gas as discussed in the next section. However, chemical and sludge disposal costs are likely to make the process uneconomical. In the brine regeneration process the high concentration (15%) of salt suggests the possibility of electrochemical treatment of the regenerant to produce nitrogen gas and chlorine.

Several attempts have been made to develop selective ion-exchange processes for nitrate removal. Dow Chemical Company is presently under contract (Contract No. 14-12-808) to develop a nitrate removal process based on the use of a porous solid absorbent containing a nitrate-selective water-immiscible extractant. The process has the advantages of liquid ion-exchange technology and the ease of operation of the granular bead resin systems.

Selective nitrate removal by ion exchange will not be feasible until new resins are synthesized with a high selectivity for nitrate over other anions present in the water. In addition, a suitable process for treating the nitrate laden regenerants must be developed.

Paper No. 7

OTHER METHODS FOR REMOVING NITROGEN

R. B. Dean

In addition to ammonia, nitrites and nitrates already discussed, nitrogen can occur in wastewaters in several other forms. Small concentrations of amino acids are leached from bacteria under anaerobic conditions. Digester supernatants contain a relatively high concentration of soluble amino acids and secondary clarifiers may release substantial quantities of amino acids if anaerobic conditions are approached in the return sludge. Amino acids are readily biodegraded and are removed in part by lime precipitation and carbon columns. Nitrilotriacetic acid (NTA) has been proposed as a detergent additive to chelate calcium ions as a partial replacement for phosphates. NTA is biodegraded in a well-operated secondary treatment plant. (Bunch and Ettinger, 1967) It is not removed by precipitation with aluminum hydroxide. The behavior of NTA toward precipitation by lime or iron salts and its behavior in the column of activated carbon containing adsorbed organic matter and a bacterial flora are now being studied at Cincinnati. NTA levels in wastewater are not expected to exceed about 2 mg/l as nitrogen.

Urea is the principal form in which nitrogen is excreted. Fortunately, urea is rapidly hydrolyzed to ammonia by bacteria in sewer lines and biological treatment plants. In the absence of biological action it would be extremely difficult to precipitate or adsorb urea from solution. Organic nitrogen is also present in particulate form both in raw sewage and in secondary effluents. In the latter case bacteria and bacterial fragments account for nearly all the suspended nitrogen. Efficient clarification and filtration are effective ways of removing this form of nitrogen. Typically, the organic nitrogen in the product from a carbon column following secondary treatment is approximately 1.0 mg N/l. This residue has not been characterized but it may be derived from bacterial action.

Soluble nitrogen compounds, with the exception of urea, are ionized in water and 85-90 percent of them can be removed by any deionizing or desalting process such as reverse osmosis, electrodialysis, ion exchange or distillation. None of these processes have a favorable selectivity for ammonium or nitrate ions and it is, therefore, necessary to remove about

90 percent of the salts in order to remove 90 percent of the soluble nitrogen compounds. Distillation is particularly unfavorable since it tends to concentrate ammonia in the distillate unless the feed is kept highly acidic. In that case, nitrous acid will be selectively distilled over (O'Connor et al., 1967). All of the general salt removal techniques are of interest only for water that is to be reused and are too expensive for consideration as treatments for wastes to be discharged to surface waters.

Chlorination of Ammonia. Ammonia reacts with chlorine stepwise according to the following equations:

(1) $Cl_2 + NH_3 \longrightarrow HCl + NH_2Cl$, monochloramine

(2) $Cl_2 + NH_2Cl \longrightarrow HCl + NHCl_2$, dichloramine

(3) $Cl_2 + NHCl_2 \longrightarrow HCl + NCl_3$, trichloramine

Monochloramine is rapidly decomposed to NH_3 and HCl by active carbon. The theoretical breakpoint equation is:

(4) $3\,Cl_2 + 2NH_3 \longrightarrow N_2 + 6HCl$

Actually $4\,Cl_2$ per $2NH_3$ or 10 mg of chlorine per mg of ammonia nitrogen is required. This means that some higher oxides of nitrogen must be produced in the reaction. A large number of possible compounds may be formed but unfortunately it is difficult to identify them analytically without very sophisticated procedures. Four moles of chlorine per mole of N_2 corresponds to 10 parts of chlorine for each part of ammonia nitrogen. The per capita excretion of ammonia nitrogen is of the order of 10 lbs. per year, equivalent to 100 lbs. of chlorine. A city of 200,000 would, therefore, require about 20 million lbs. or 10,000 tons of chlorine each year. Transport and handling of this amount of chlorine would be hazardous. Also, the use of chlorine would produce an equivalent weight of hydrochloric acid, which would exceed the buffer capacity of most waste waters, and would produce the highly odorous NCl_3. While hydrochloric acid would be effectively neutralized by discharge into the ocean and most surface waters, it could be a serious pollutant and discharge pipes would have to be made of special acid resistant materials. Sodium or calcium hypochlorite can be substituted for chlorine in equations (1-4).

$$(5)\quad 3NaOCl + 2NH_3 \longrightarrow N_2 + NaCl + 3H_2O$$

No excess hydrochloric acid is produced by the use of sodium hypochlorite, but again it will be necessary to use four moles of hypochlorite per mole of nitrogen or equivalent. Hypochlorite is more expensive than chlorine but is so much safer to use that New York City, Chicago, and other large cities have converted to aqueous hypochlorite for chlorination of water and treated wastes.

Chlorine is manufactured electrolytically according to equation (3):

$$(6)\quad 2NaCl + 2H_2O \longrightarrow H_2 + Cl_2 + 2NaOH$$

Sodium hypochlorite is made by the reaction of chlorine and NaOH. If this is done in the electrolytic cell, equation (4) results.

$$(7)\quad NaCl + H_2O \longrightarrow H_2 + NaOCl$$

Electrochemical production of chlorine or sodium hypochlorite has been practiced for sewage chlorination and treatment for years (Marson, 1967). It has never been optimized. Some source of brine is necessary to raise the conductivity of the water; therefore, electrochemical production of chlorine or hypochlorite has only been attractive for seacoast communities. Breakpoint chlorine treatment, of course, also disinfects as well as destroys ammonia. However, the addition of 200-300 mg/l of chloride might not be acceptable in inland waters.

Just as ammonia can be oxidized to nitrogen, so can nitrate be reduced to nitrogen. In acid media, nitrous acid derived from nitrites reacts with ammonia and amines to liberate nitrogen gas.

$$(8)\quad HNO_2 + RNH_2 \longrightarrow ROH + N_2 + H_2O$$

This reaction, called the Van Slyke reaction, does not proceed in neutral waters. (Brezonik and Lee, 1966) Chemical reduction of nitrates in dilute solution has been accomplished using ferrous sulfate to produce nitrogen gas and some nitrous oxide, N_2O. (Gunderloy et al., 1968) The by-product is a voluminous sludge of ferrous-ferric hydroxide. Some of the sulfate is incorporated in the precipitate but a large part of the sulfate remains in the waste. This process is being studied under contract and at present appears to have marginal value for special situations.

Costs of any of the processes discussed in this section are probably higher than costs for the more promising processes. The only cost that can be calculated direct with some confidence is the cost for chlorine to oxidize ammonia. Assuming 20 parts of ammonia nitrogen in a secondary effluent, 200 parts of chlorine would be required for breakpoint chlorination. This is equivalent to 1.5 lbs. of chlorine per 1000 gallons or about 6¢/1000 gal. To this must be added the cost of handling the corrosive hydrochloric acid produced. Sodium hypochlorite may cost twice as much as chlorine but associated costs are greatly reduced. The cost of electrically generated chlorine or hypochlorite will depend on the cost of electrodes and the possibility of using low-cost off-peak power.

SALT CREEK TWO STAGE NITRIFICATION PLANT
METROPOLITAN SANITARY DISTRICT OF GREATER CHICAGO

by

Richard L. Woodward, Associate
Camp, Dresser & McKee

The proposed Salt Creek Water Reclamation plant of the Metropolitan Sanitary District of Greater Chicago will serve an area of some 49 square miles in the northwestern part of Cook County about 20 to 25 miles northwest of downtown Chicago. It is primarily a residential area with some light industry. A portion of the area is served by combined sewers. At present, the area is tributary to the North Side or West-Southwest treatment plants but because of the rapid growth of this portion of metropolitan Chicago a decision was reached following a 1965 study by Greeley and Hansen to construct several treatment plants to serve the area rather than to build relief interceptors and to expand the existing plants.

Salt Creek is a tributary of the Des Plaines River. It rises in northwestern Cook County and flows across the northeastern part of Du Page County before reentering Cook County and joining the Des Plaines River. The outfall sewer from the Salt Creek treatment plant will discharge to Salt Creek a short distance above the Cook-Du Page county line and a short distance downstream from a proposed reservoir which is to be built in the Ned Brown Forest Preserve. During the summer months there will be virtually no dilution available for the effluent from the treatment plant. Through much of its length Salt Creek flows through park and forest preserve lands which are used for recreation by large numbers of people. Although the stream is not used to any appreciable extent for body contact recreation it is important that it be maintained as an attractive part of the landscape.

The Illinois Sanitary Water Board has established water quality standards and effluent standards which require an effluent of quite high quality from the Salt Creek plant. Some of the most pertinent of these are:

	Standards	Influent Wastewater
BOD	4 mg/l	95 mg/l
Suspended Solids	5 mg/l	188 mg/l
Ammonia Nitrogen	2.5 mg/l	20 mg/l
Nitrate Nitrogen	45 mg/l	--
Fecal Coliforms	2,000 per 100 ml	--

There is not at present any effluent or stream standard for phosphates in this area.

The wastewater to be treated at the Salt Creek plant is comparatively dilute. A sampling program of one week's duration of the principal sewer which will be tributary to the plant showed the average characteristics summarized in the table.

Treatment Processes Considered

The high effluent quality demanded requires that the treatment process remove most of the carbonaceous and nitrogenous oxygen demanding materials. The two principal processes which had been developed at the time of our studies (late 1967) for removing ammonia were air stripping, which had been studied at Lake Tahoe and nitrification which had been studied by a great many investigators.

Ammonia stripping was considered but was found to be more costly and less reliable than nitrification. The high alkalinity and hardness of this wastewater would have required very large quantities of lime and produced correspondingly large quantities of sludge to be processed. At the low temperatures which prevail in the Chicago area in the winter the process would be quite inefficient in ammonia removal and would probably have to be discontinued because of icing and other operating difficulties.

The nitrification process in activated sludge systems is fairly well understood now as a result of the work of a number of investigators both here and abroad. Downing and his colleagues at the Water Pollution Research Laboratory in England have contributed most significantly to this understanding. The basic requirement for maintaining a nitrifying flora in an activated sludge system is that the overall rate of sludge growth (and hence of sludge wasting) be no greater than the rate of growth of the nitrifying organisms. In a conventional system this requires relatively long cell residence times in order that much of the sludge produced as a result of the activity of the heterotrophs in removing carbonaceous material will be aerobically digested. However, for one reason or another, many plants which provide long cell residence times appear to be rather erratic in nitrifying. Just why this is so remains puzzling.

Barth and his colleagues at the Robert A. Taft Sanitary Engineering Center have worked with a two-stage activated sludge system which appeared to offer a number of advantages over the conventional one-stage systems where reliable nitrification is called for. In this system, the first aeration stage is devoted to removal of most of the carbonaceous material and the second stage largely to nitrification. By maintaining these two separate specialized sludges it is possible to achieve such long residence times in the second stage that nitrification will proceed even at relatively low temperatures without providing as long aeration times as would be needed in a conventional system.

After some comparative cost studies this two-stage system was recommended to the Sanitary District and was approved by them. To achieve the very low suspended solids and BOD concentrations called for, filters were provided following the second-stage clarifiers. Chlorination will be provided either before or after the filters to meet the coliform standard of the Illinois Sanitary Water Board.

Nutrient Removal

At present, the prescribed water quality standards do not require removal of either nitrogen or phosphorus, but this type of plant has the capability of providing for nutrient removal without appreciable additional capital cost. Such provisions have been made in the design of the plant.

Phosphate removal can be accomplished by adding an iron or aluminum salt to the first aeration stage. This has been studied by Barth and his colleagues as well as by a number of other investigators. At the Salt Creek plant provision has been made for such chemical additions.

Nitrogen removal by denitrification can be accomplished in the filters which follow the second-stage clarifiers by providing a carbon source to permit the development of a denitrifying flora in the filters. This process has been investigated at Pomona, California by the Federal Water Quality Administration and the Los Angeles County Sanitation District. Provisions for feeding methanol as a carbon source are included in the Salt Creek design.

It is not known to what extent nutrient removal will be practiced at Salt Creek, but it was relatively inexpensive to provide such capability.

Pilot Plant Studies

The Sanitary District was anxious to proceed immediately to final design of the Salt Creek plant after approval of our report in early 1968. The deadline for completion of final plans and specifications was December, 1969. Provisions were made to convert a one million gallon per day sewage treatment plant which the District operates at Hazelcrest, Illinois in southern Cook County to the type of process contemplated for Salt Creek. This was done and the plant was operated during part of 1969. It was not possible to develop a nitrifying flora until the sewage temperature reached about 60° F, but thereafter the plant operated smoothly with almost complete nitrification even after cold weather had reduced sewage temperatures to the 40's.

The nutrient removal facilities of the plant were tested to only a limited extent. No problems were experienced with phosphate removal by feeding sodium aluminate. The denitrification step encountered difficulties which were related more to equipment provided in the pilot plant than to the process. These should not be a problem in the full scale plants.

Plant Design

The design calls for the plant to handle an average flow of 30 mgd during dry weather with 50 mgd on maximum dry weather days. During wet weather 50 mgd will be given complete treatment. Wet weather flows above 50 mgd will receive sedimentation and chlorination. After coarse racks, pumping, screening and grit removal the sewage will flow to the first aeration stage which will provide three hours detention at 50 mgd. Sedimentation will be in circular clarifiers with an overflow rate of 1,450 gallons per day per square foot based on maximum day flow plus 30 percent sludge return. The second-stage clarifiers and sedimentation tanks will be the same as the first stage. Filters will have three feet of anthracite and sand media and will be designed to operate at six gallons per minute per square foot.

Sludge will be thickened in air flotation thickeners, digested and pumped to a land reclamation site. The Salt Creek plant will provide digestion facilities for sludge from another plant to be built in this part of the Sanitary District.

Chemical feeding facilities will permit feeding alum to the first aeration stage and also to the influent to the filters. Polyelectrolytes can also be fed as filter aids to the filter influent and to assist in sludge thickening. Methanol can be added to the filter influent if denitrification is desired. Chlorine can be added at a number of points in the system to deal with odor problems or to disinfect.

The plant is designed as the first stage of construction to be expanded ultimately to three times its initial size.

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ADVANCED WASTE TREATMENT AND WATER REUSE SYMPOSIUM

PHOSPHORUS REMOVAL IN CONVENTIONAL TREATMENT

Edwin F. Barth
Advanced Waste Treatment Research Laboratory
U.S. Department of the Interior
Federal Water Quality Administration
Ohio Basin Region
Cincinnati, Ohio

GPO 821-871-12

PHOSPHORUS REMOVAL IN CONVENTIONAL TREATMENT

Only 60% of the population of the United States provides some type of treatment for its wastewater. Most conventional treatment systems are biological processes and these are generally inefficient in the removal of phosphorus. Annually in the United States about 400×10^6 pounds of phosphorus are discharged to receiving waters from these treatment plants. Another 200×10^6 pounds of phosphorus are discharged from individual home units or other unsewered sources.

Of the treatment processes employed in conventional systems the activated sludge process is most efficient in phosphorus removal because the efficiency of removal is directly related to solids production. Several activated sludge plants in various parts of the country have reported very efficient removals. Evaluation has indicated the operating parameters, at these particular sites, which influence phosphorus removal. In most cases these removals can be ascribed to interaction of the chemical characteristics of the carriage water and the biological process.

One key observation at these sites has been the fact that in a conventional system digester supernatant recycle has a great influence on phosphorus removal. Anaerobic digestion solubilizes the solids previously removed in the main stream and recycle of the supernatant adds large quantities of soluble phosphorus, organic matter, and nitrogenous material into the biological process.

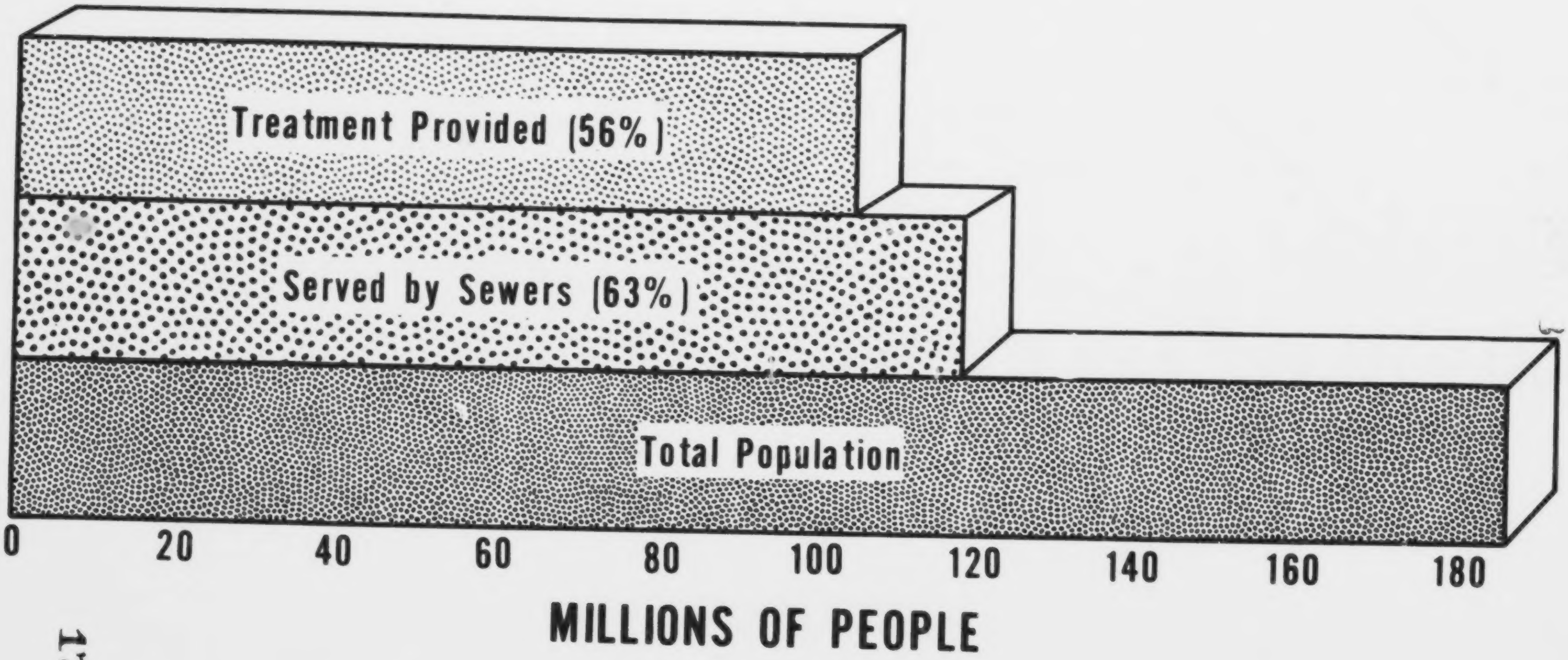
Aside from this nutrient loading aspect, digester supernatant introduces a high organic load and much colloidal material. A chemical-physical process has been developed which can effectively treat supernatant. Application of this process at treatment plants which are of marginal efficiency can up-grade the quality of the final effluent.

At those plants which are marginal we can do several things to increase nutrient removal efficiency and generally produce an overall improvement in effluent quality. Most marginal performance is due to four causes. These are hydraulic overload, organic overload, poor liquid-solids separation, and operator skill.

There is a large amount of capital investment in existing conventional secondary treatment facilities, and provision for efficient performance of these facilities can go a long way in meeting our water quality goals. Many of these units can be modified to improve removal of specific pollutants, such as phosphorus; or serve a useful function as a first step in an advanced process. Much of our past effort has been in the direction of reducing the cost of treatment, not in the direction of improved treatment. In many areas we have low cost treatment and low efficiency.

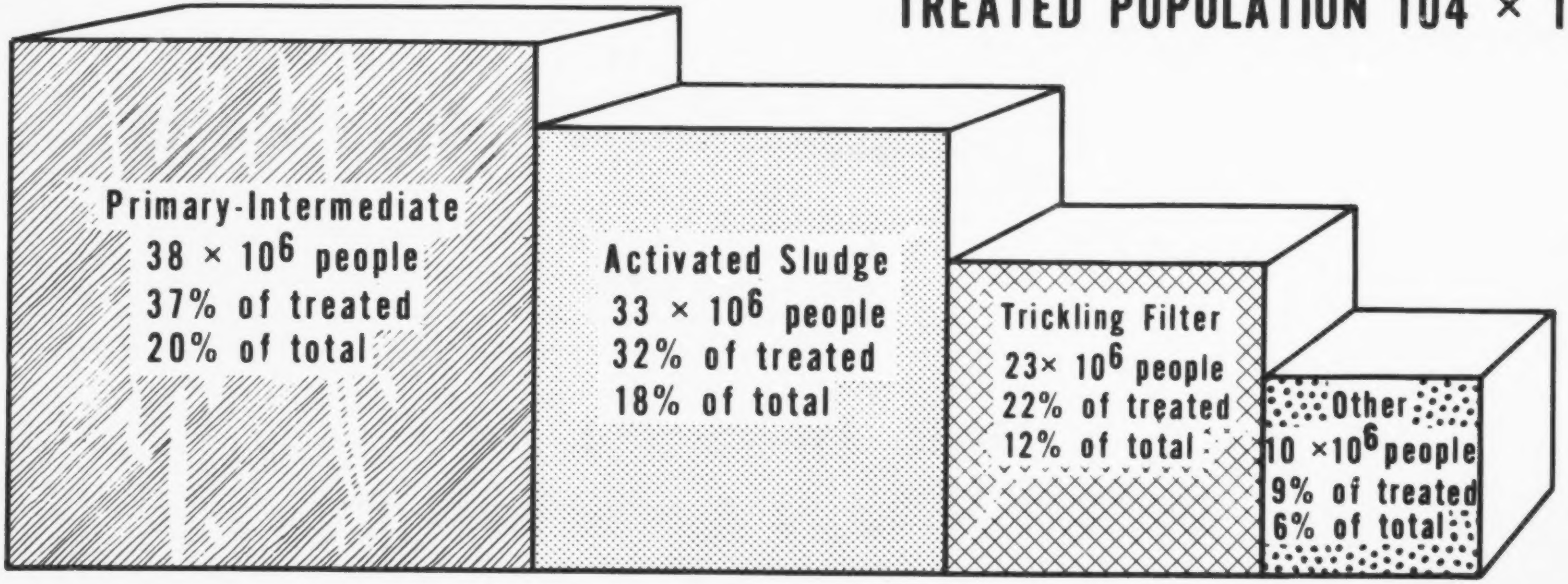
Attached are the group of tables and figures that will be used to present this subject.

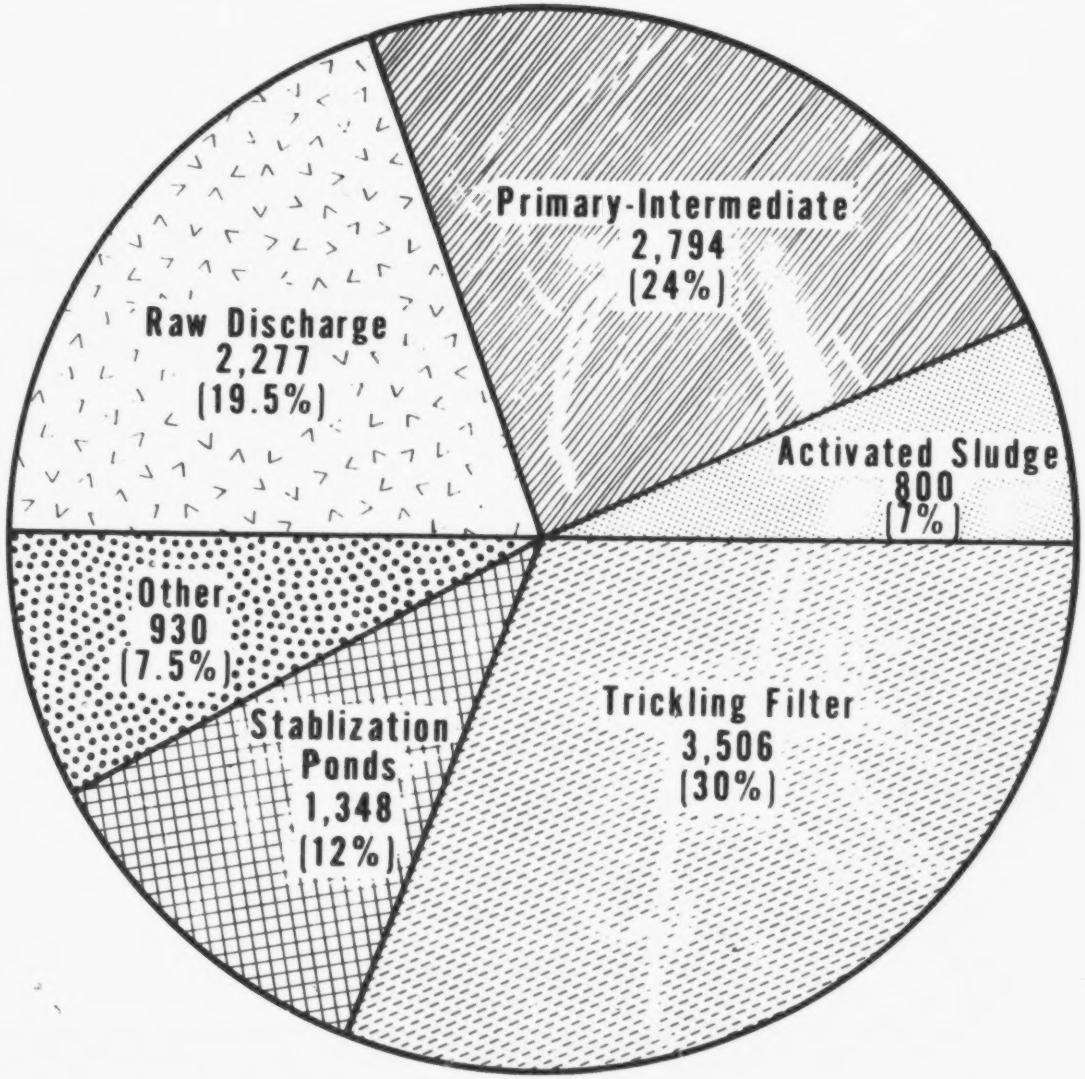
POPULATION WITH WASTE SYSTEMS
PHS SURVEY 1962



TREATMENT CLASSIFICATION BY POPULATION
PHS SURVEY 1962

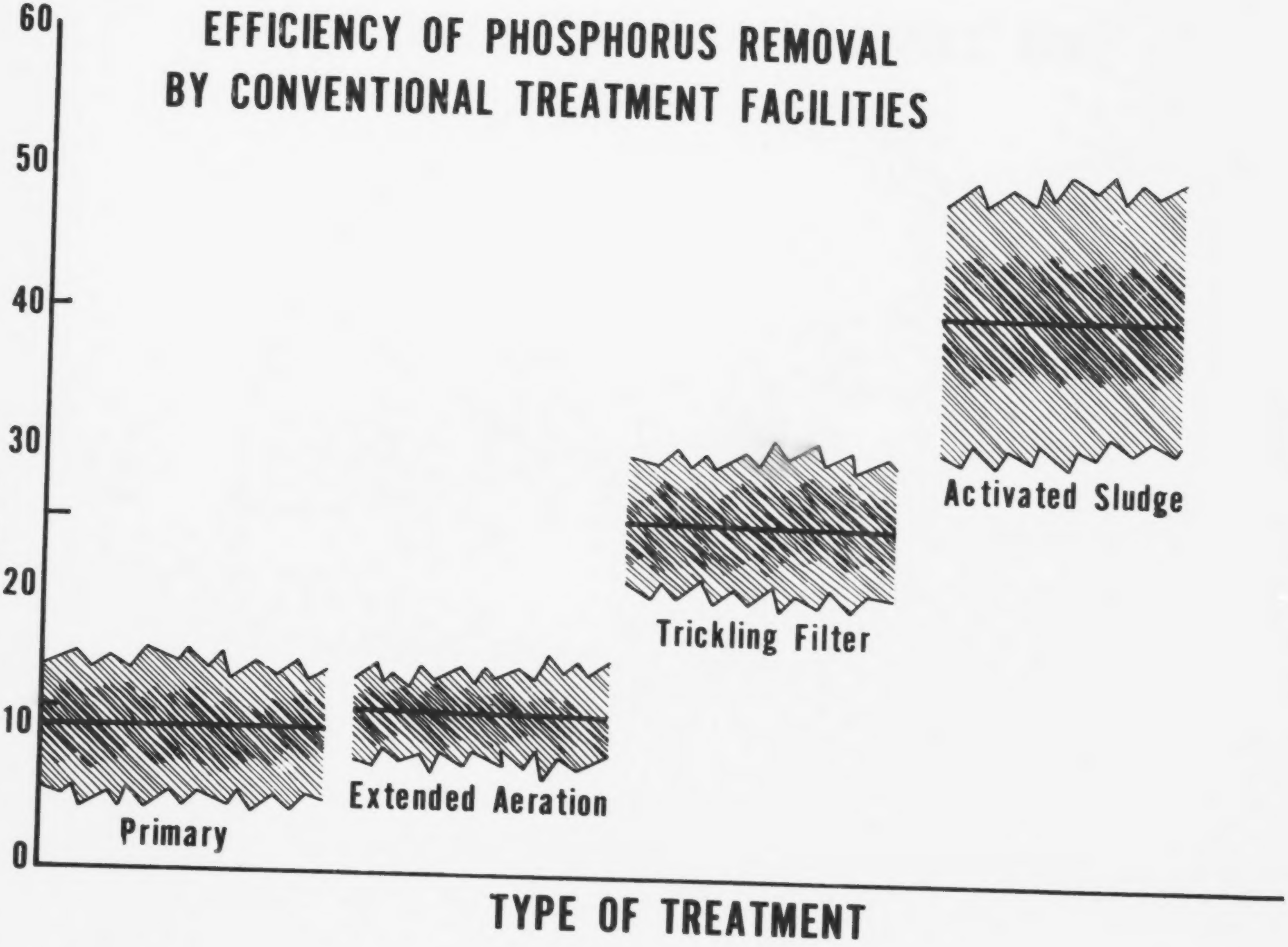
TOTAL POPULATION 186 × 10^6
TREATED POPULATION 104 × 10^6





TOTAL FACILITIES 11,655 PHS SURVEY, 1962

EFFICIENCY OF PHOSPHORUS REMOVAL BY CONVENTIONAL TREATMENT FACILITIES



ESTIMATE OF TOTAL AMOUNT OF PHOSPHORUS DISCHARGED TO WATER SYSTEM

Per capita contribution of phosphorus is 4 pounds per person per year

Population x per capita x efficiency	Pounds discharged
Not served by any treatment $70 \times 10^6 \times 4 \times 0.00$	280×10^6
Served by primary treatment only $40 \times 10^6 \times 4 \times 0.10$	154×10^6
Served by trickling filter or extended aeration $30 \times 10^6 \times 4 \times 0.25$	90×10^6
Served by activated sludge $40 \times 10^6 \times 4 \times 0.40$	96×10^6
Total pounds of phosphorus discharged to water system per year	610×10^6

PARAMETERS DEEMED NECESSARY FOR EFFICIENT BIOLOGICAL REMOVAL OF PHOSPHORUS

1. Eliminate Digester Supernatant Recycle
2. Aeration Time and Rate of Air Supply
3. Concentration of Mixed Liquor Suspended Solids
4. Concentration of Dissolved Oxygen in Aerator
5. Period of Sludge Retention in Secondary Clarifier
6. Additional Benefit Possible by Stripping of Phosphorus from Return Sludge
    - a) Anaerobic Stripping
    - b) Acid Stripping

Summary of Supernatant Data (6)

Analysis	80% Confidence	Median	Mean
Bicarbonate Alkalinity (mg/l as $CaCO_3$)	850 - 2,950	1,580	1681
5-Day BOD (mg/l)	500 - 4,260	910	1401
Total Phosphate (mg/l as PO_4)	46 - 380	132	211
(mg/l as P)	15 - 124	43	69
Ammonia Nitrogen (mg/l)	100 - 710	405	413
pH	6.6 - 8.0	7.3	7.3

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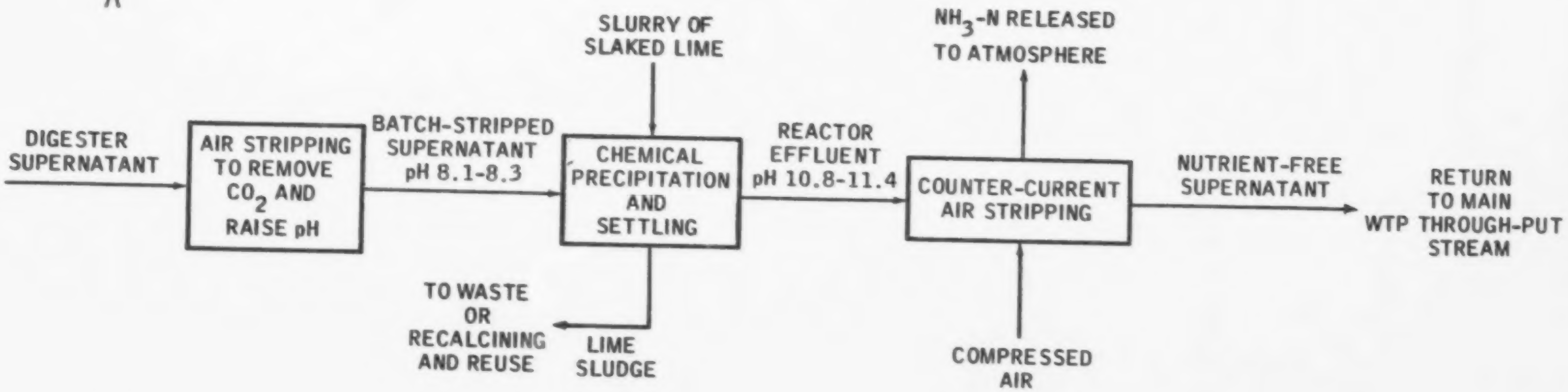
(6) Masselli, Joseph W. et.al. "The Effect of Industrial Wastes on Sewage Treatment". New England Water Pollution Control Commission, Boston, Massachusetts (1965).

SUPERNATANT BENEFICIATION PILOT PLANT TREATMENT SEQUENCE

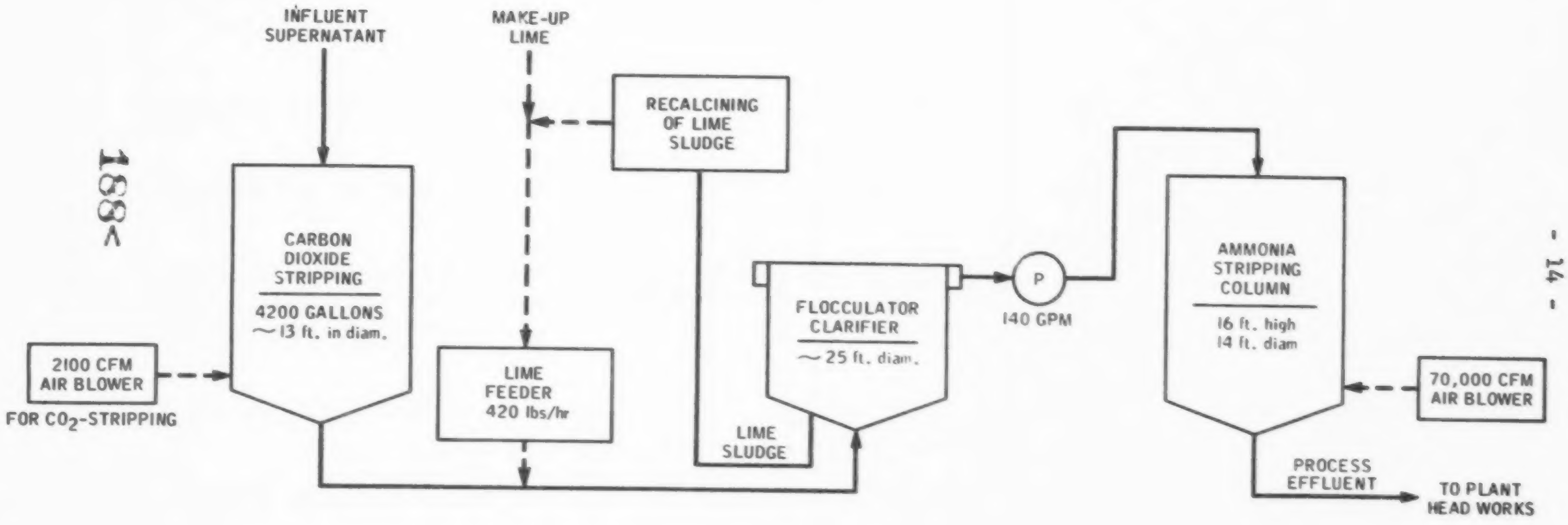

DIGESTER SUPERNATANT
AIR STRIPPING TO REMOVE CO_2 AND RAISE pH
BATCH-STRIPPED SUPERNATANT pH 8.1-8.3
SLURRY OF SLAKED LIME
CHEMICAL PRECIPITATION AND SETTLING
LIME SLUDGE
TO WASTE OR RECALCINING AND REUSE
REACTOR EFFLUENT pH 10.8-11.4
NH_3-N RELEASED TO ATMOSPHERE
COUNTER-CURRENT AIR STRIPPING
COMPRESSED AIR
NUTRIENT-FREE SUPERNATANT
RETURN TO MAIN WTP THROUGH-PUT STREAM

OVERALL PILOT PLANT PERFORMANCE
(RUN NOS. 1 – 13)

	AVERAGE INFLUENT (PPM)	AVERAGE EFFLUENT (PPM)	AVERAGE REMOVAL
Overall Ortho-P removal	65	5	92 %
Overall Total Phosphorus removal	138	22.5	84.5%
Overall NH_3-N removal	812	175	79%
NH_3-N removal through first stripping tower	763	330	57.5%
NH_3-N removal through both stripping towers	763	175	77.5%
Overall TOC removal	1202	590	51%
Overall Volatile S.S. removal	1679	426	71%
Overall Suspended Solids removal	2262	647	69%

March 1970

TYPICAL
SUPERNATANT BENEFICIATION FACILITIES FOR 50 MGD PLANT*


INFLUENT SUPERNATANT
MAKE-UP LIME
RECALCINING OF LIME SLUDGE
2100 CFM AIR BLOWER
FOR CO2-STRIPPING
CARBON DIOXIDE STRIPPING
4200 GALLONS
~13 ft. in diam.
LIME FEEDER
420 lbs/hr
LIME SLUDGE
FLOCCULATOR CLARIFIER
~25 ft. diam.
P
140 GPM
AMMONIA STRIPPING COLUMN
16 ft. high
14 ft. diam
70,000 CFM AIR BLOWER
PROCESS EFFLUENT
TO PLANT HEAD WORKS

*PRODUCING IRVINGTON-TYPE SUPERNATANT

CONTROL OF HYDRAULIC OVERLOAD

1. FIND AND REDUCE NEEDLESS SOURCES
2. CONSIDER USING LARGE INTERCEPTORS AS HOLDING TANKS
3. CONSTRUCT SURGE TANK
   - FANCY
   - PLAIN
4. ENLARGE PLANT

CONTROL OF ORGANIC OVERLOAD

1. SURGE TANK TO SMOOTH LOAD
2. HAVE INDUSTRY PROGRAM LOAD
3. POLYMERS DURING BULKING
4. PROGRAM DIGESTER SUPERNATANT RECYCLE OR ELIMINATE
5. INSTALL ROUGHING FILTER-IN LINE OR LOOP OPERATION
6. CONSIDER TWO-STAGE TREATMENT

IMPROVE FINAL SOLIDS SEPARATION

1. CONTROL RECYCLE
2. POLYMERS OR WEIGHTING AGENTS
3. LAGOON
4. COARSE-SAND FILTRATION
5. INCREASE SETTLER CAPACITY
6. ADVANCED PROCESSES

ADVANCED WASTE TREATMENT AND WATER REUSE SYMPOSIUM

COMBINED BIOLOGICAL-CHEMICAL TREATMENT FOR CONTROL OF PHOSPHORUS

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COMBINED BIOLOGICAL-CHEMICAL TREATMENT FOR CONTROL OF PHOSPHORUS

Richard C. Brenner

INTRODUCTION

The sanitary engineer has three basic options available today in designing waste treatment systems to remove phosphorus. As shown in Figure 1, the three options are straight biological removal, straight chemical precipitation, and combined biological-chemical removal.

Biological removal is dependent on manipulating aeration system operating parameters to maximize cell synthesis and so-called luxury phosphorus uptake. Overall removal efficiency is improved by breaking the digester recycle and treating supernatant separately. To date, however, luxury phosphorus uptake has not proven to be a predictable phenomenon which can be effectively applied to activated sludge plants across the board. For reliable phosphorus removal, one of the other two options should be selected.

The major difference in straight chemical precipitation and combined biological-chemical removal is the point of chemical addition. Strict chemical precipitation schemes add the chemicals either prior to the biological unit in the primary settler or after the biological unit in a tertiary step using biological effluent as feed. Other examples of straight chemical applications are the new physical-chemical plants which do not include a biological phase. Combined biological-chemical removal utilizes direct dosing of mineral salts to the biological unit itself.

PHOSPHROUS PRECIPITANTS

The available chemicals which we know today will effectively precipitate phosphorus from solution are listed in Figure 2. They include aluminum and iron salts and lime. Precipitation with metal salts is based on the low solubility of metallic phosphorus compounds at typical wastewater pH ranges. The required dose of the metal ion is stoichiometrially related to the incoming phosphorus concentration. Precipitation with lime is independent of phosphorus concentration; it is dependent on pH and bicarbonate alkalinity. The rationale for lime precipitation is the very low solubility of hydroxyapatite at elevated pH.

All of the above chemicals are applicable to strict chemical precipitation systems. Only the metallic salts, however, are effective and economically realistic when dosed directly to a biological treatment unit. Excessive doses of lime are needed to overcome the natural buffering action of biological systems.

When metallic salts are added to the primary settler to remove phosphorus, small doses of anionic polymers are also recommended to improve flocculation and settleability. If the polymer is not added, much of the insolubilized phosphorus particles may carry over into the next step of the treatment sequence. It may further be necessary in certain instances to supply additional alkalinity in the form of lime or caustic soda to the primary settler to achieve effective flocculation and solids capture. This is particularly true with ferrous chemicals.

The presence of naturally occurring polymeric materials in activated sludge eliminate the necessity of commercial polymer aids in combined biological-chemical applications. The biological floc serves as the coagulant mechanism to capture precipitated phosphorus particles.

BASIC PRINCIPLES OF COMBINED BIOLOGICAL-CHEMICAL REMOVAL (MINERAL ADDITION)

The practice of adding chemicals to activated sludge plants to remove phosphorus will be referred to hereinafter as mineral addition. Precipitation of phosphorus takes place in the aeration tank. The metallic-rich sludge which is formed is continuously separated in the secondary settler and recycled to the head of the aerator. Adsorption and flocculation of phosphate precipitates is enhanced by the presence of the combined biological-chemical sludge and biologically generated polymeric materials. Precipitated phosphorus is removed from the process in the waste activated sludge. The waste sludge from a mineral addition process can be thickened and disposed of with the same techniques used for conventional biological sludges. Excepting that which is incorporated in cells to satisfy metabolic needs, the phosphorus removed in the sludge is chemically bound and is not released during aerobic or anaerobic digestion.

Figure 3 indicates alternative dosing points for mineral addition. The minerals can be added with relatively equal phosphorus removal expectancy to the primary effluent, a flash mix tank just prior to final sedimentation, the recycle sludge, or at any point in the aeration tank. Individual plant configurations and a few trial and error runs will usually dictate the most preferred dosing point for a particular situation.

ADVANTAGES OF MINERAL ADDITION

The mineral addition technique possesses several inherent advantages for activated sludge plants. These advantages are summarized in Figure 4 and include minimum capital expenditure, reduced chemical requirement, improved mixed liquor settling and the resultant capability of carrying higher mixed liquor volatile suspended solids loadings in the aerator, and a large reduction in phosphorus recycle back to the main flow stream via digester supernatant.

Capital expenditures are limited to chemical feed equipment, storage facilities, and dosing control instrumentation. Mixing is accomplished in the aeration tank and phosphorus precipitates are removed in the existing settler.

Reduced chemical requirements can be traced to biological-chemical interaction and the large floc surface area available for adsorption. As shown in Figure 5, the alum requirement for equivalent phosphorus removal using the mineral addition approach is only about 75% of that for a strictly tertiary application.

In comparing mineral requirements for dosing to the aerator versus dosing to the primary, there is the further consideration of the form of the phosphorus. Inorganic phosphorus in raw sewage is normally divided about equally between polyphosphate and orthophosphate. Polyphosphate is rapidly hydrolyzed to orthophosphate by the enzymatic action of activated sludge. Precipitation of phosphorus is much more efficient in the ortho form than in the poly form. Thus, less mineral supplement is needed in the aerator than for an equivalent removal in the primary.

Figure 6 shows the results of a pilot plant experiment conducted several years ago at the Advanced Waste Treatment Research Laboratory (AWTRL) in Cincinnati (1). Sodium aluminate was added to the pilot plant after an initial base run period. The improvement in sludge density index effected by mineral addition is readily apparent. This made possible an approximate two-fold increase in the concentration of mixed liquor volatile suspended solids, thereby nearly halving the organic loading to the system. In general, the greater the bulking tendency of a sludge, the more marked will be the improvement in settling characteristics with mineral addition.

The ultimate fate of phosphorus with mineral addition is also evident in Figure 6 (1). The phosphorus concentration in digested sludge in the above experiments rose by 235% after aluminum addition while the concentration in the supernatant fell to a third of its former level. Phosphorus precipitated by the aluminum ion was not released during anaerobic digestion. That which did remain in the supernatant was solubilized during digestion from the primary sludge which received no chemical dosing and from non-chemically bound phosphorus used to satisfy cell matabolic needs.

LIMITATIONS OF MINERAL ADDITION

It is important to understand the limitations of a process to use it properly. The mineral addition process for activated slusge plants is highly efficient in removing the bulk of phosphorus from wastewaters.

Effluent concentrations of 1-2 mg/l as P are readily achievable. The difference between 1 and 2 mg/l is largely a function of suspended solids removal. If final clarification is followed by sand filtration, effluent phosphorus can be further lowered to 0.5-1.0 mg/l as P. It is difficult to obtain residual phosphorus levels below 0.5 mg/l with mineral addition due to the solubility products of aluminum and iron phosphates. If residual phosphorus levels below 0.5 mg/l are mandatory, tertiary lime precipitation is recommended.

COMPARISON OF ALUMINUM AND IRON PRECIPITANTS

Two aluminum compounds have proved satisfactory for mineral addition. They are sodium aluminate and aluminum sulfate (alum). Four iron compounds have been used by various researchers including ferric chloride, ferrous chloride, ferric sulfate, and ferrous sulfate.

Three inherent factors favor use of the aluminum products over the iron products. First, iron forms colored compounds; aluminum compounds are white. Effluent discoloration is a greater possibility with iron than with aluminum. Second, iron undergoes oxidation-reduction reactions; aluminum does not. Therefore, the possibility does exist that iron phosphate sludge could be reduced during digestion and the phosphorus subsequently resolubilized. Reduction is not possible with aluminum. Third, all of the iron compounds are highly acidic, whereas both acidic and basic forms are available with aluminum. Aluminum thus offers a choice in selecting the proper chemical to match wastewater alkalinity.

Favoring iron is the fact that commercial grade iron compounds are slightly cheaper than commercial grade aluminum compounds for equivalent phosphorus removal. Also, iron compounds are more frequently available as impure waste streams, e.g. pickle liquor. These impure waste streams can usually be purchased for the cost of trucking alone, thereby significantly lowering the cost of phosphorus removal.

COMPARISON OF ALUM AND SODIUM ALUMINATE

The caracteristics of the two aluminum salts are summarized in Figure 7. Alum is highly acidic, uses up alkalinity reserve, and lowers pH. Sodium aluminate is highly basic, acts as a natural buffer, and causes pH to elevate slightly.

Both compounds are available in liquid and powder form. The alum forms contain a higher percentage of water than their aluminate counterparts. Alum is produced in more than fifty locations around the country. There are only three major aluminate production centers. Delivery of liquid alum is limited to approximately 100-125 miles. At distances greater than this, powdered alum becomes more economical.

Both liquid alum and aluminate are more easily handled and metered into a treatment process than the powder forms. No dissolving equipment is required for the liquids. However, temperature is a factor to consider in both transport and storage of the liquid forms. Liquid alum crystallizes at 4°F; liquid aluminate at -20°F. Both materials, particularly liquid aluminate, become increasingly viscous as temperature drops. Indoor storage is preferred. Insulated tanks and steam tracing will be necessary in colder climates for outdoor storage.

SIDE EFFECTS OF MINERAL ADDITION

Regardless of the precipitant used, some extraneous and unusuable ion is added to the water. These extraneous ions are a form of pollution themselves. In the cases of alum, ferric sulfate, and ferrous sulfate, sulfate is the extraneous ion which remains in solution. For ferric and ferrous chloride, the extraneous ion is chloride. Sodium is the extraneous ion for sodium aluminate. The quantities of extraneous ion added per pound of metal ion are shown in Figure 8. The increase in total dissolved solids for equivalent phosphorus removal using aluminate is six times less than for alum and 3-6 times less than for the various iron compounds.

MATCHING OF PROPER CHEMICAL TO WASTEWATER ALKALINITY

Two examples will be utilized below to illustrate the importance of considering wastewater alkalinity before selecting the mineral additive.

The first example is taken from data collected during experiments at the AWTRL in Cincinnati in which alum and aluminate were dosed separately to two parallel and identical activated sludge pilot plants (2). Doses to the two plants as mg/l of aluminum ion were equivalent. The same wastewater with a moderate alkalinity varying from 100-200 mg/l as $CaCO_3$ served as the feed to both plants. As seen in the sequence of Figures 9 through 12, the performance of the alum fed unit was much more erratic than that of the aluminate fed unit. This can be traced to the nearly constant decrease in alkalinity of 100 mg/l as $CaCO_3$ exhibited in the alum fed unit. When the alkalinity of the primary effluent entering the alum fed aerator was down near 100 mg/l, the alum caused the alkalinity to approach zero with a concurrent substantial drop in pH (day 16). The microbial population, not being acclimated to the lower pH, suffered through a several day recovery period during which effluent turbidity and phosphorus rose significantly. Alkalinity and pH of the aluminate fed unit were relatively unaffected and performance was consistent throughout the study. Overall performance of the two parallel pilot units for various treatment parameters is summarized in Figure 13. In this case, aluminate is clearly the preferred additive.

In a contrasting study at the joint FWQA-Los Angeles County Sanitary District pilot plant in Pomona, California, the opposite situation was observed (3). The wastewater at this site has a fairly high alkalinity of 300-400 mg/l as $CaCO_3$. At typical phosphorus concentrations, the required alum dose is far short of that which would reduce alkalinity to low levels. Aluminate, however, being basic, tends to increase hydroxide alkalinity and pH. Comparative unfiltered effluent phosphorus values in Figure 14 indicate superior performance with alum at equivalent aluminum ion doses. In this instance, alum, or any of the iron compounds which are also acidic, is a better choice than aluminate.

From these two examples, it can be postulated that aluminate is better suited to wastewaters with low to moderately low alkalinity while alum and iron compounds are better suited to wastewaters with moderately high to high alkalinity.

REQUIRED MINERAL DOSES - PILOT PLANT STUDY

In the same Cincinnati study (2) mentioned previously in which alum and aluminate were compared, % total phosphorus removed across the secondary was plotted against the dosing ratio of aluminum ion to phosphorus (Al/P). Discarding the data from the alum fed unit on those days when the unit was upset, alum and aluminate exhibited equal capability to remove phosphorus.

The following information can be extracted from the curve plotted in Figure 15. At an Al/P dose of 1/1, by weight (mole ratio of 1.15/1), 80% phosphorus removal can be expected. Removal expectancy is increased to 90% at an Al/P dose of 1.2/1, by weight (mole ratio of 1.4/1). Above 1.2/1, there is a diminishing return in incremental phosphorus removed. Maximum removal of 93% total phosphorus was achieved at Al/P doses of 1.4/1 to 1.5/1, by weight. Increasing the dose further only succeeded in decreasing overall removal. One hypothesis for this phenomenon is that stable aluminum phosphate colloids were formed with the excessive Al doses that resisted settling.

Assuming equivalent phosphorus removal at mole ratios of 1.15/1 and 1.4/1, the required Fe/P doses to achieve 80% and 90% removal are 2.1/1 and 2.5/1, by weight, respectively.

FIELD STUDIES AT ACTIVATED SLUDGE PLANTS

Two full scale field studies can serve to demonstrate the applicability of the mineral addition process to actual activated sludge plants.

The first was conducted by AWTRL staff personnel at the 1 mgd Xenia, Ohio Little Miami Plant in the summer of 1967 (4). In this six-week study, sodium aluminate was added directly to the head of the aerator.

Figure 16 shows that phosphorus removal approached 90% at Al/P doses of 1.5/1 to 1.8/1 by weight. The temporary dosing system utilized did not always properly match Al to P with the result that overall phosphorus removal suffered slightly. Improvement in activated sludge settling and concentration was noted throughout.

The second study was conducted in 1969 under an FWQA research grant to the University of Texas, Medical Branch, on Texas City, Texas Plant No. 2 (5). Waste pickle liquor ($FeSO_4$) was dosed directly to raw sewage before primary sedimentation. Since no polymer was added, the primary settler showed negligible phosphorus removal. Insolubilized phosphorus particles carried through to the aeration tank where overall removal was excellent. As indicated in Figure 17, an effluent phosphorus level of approximately 1 mg/l as P was reliably achieved at an Fe/P mole ratio of 1.4/1 to 1.5/1. Upon cessation of ferrous addition on June 13 (Figure 18), effluent phosphorus increased dramatically to a background level of 9-10 mg/l as P. Equally effective results were obtained throughout the dosing program even though mixed liquor suspended solids concentrations varied from 2000-4200 mg/l. The Texas City study demonstrated that insolubilized phosphorus has little affinity for settling in the primary without polymer. When primary sedimentation is followed with activated sludge treatment, dosing of the mineral to raw sewage or primary effluent appears to make little difference in overall phosphorus removal efficiency.

DIRECT DOSING OF A TRICKLING FILTER

To ascertain the applicability of mineral addition to trickling filters, one of three Fairborn, Ohio trickling filters was dosed directly with sodium aluminate for a four-week period in the summer of 1968 (6). Total flow to the plant of 3 mgd was split equally between the three filters. One of the undosed filters was used as a control. The results are given in Figure 19. Overall phosphorus removal with the dosed filter averaged 64% compared to 17% for the control filter. Performance was otherwise unaffected.

It can be concluded that direct mineral dosing is considerably less effective for a trickling filter than for activated sludge. The apparent reasons are insufficient mixing, reduced residence time, and the absence of freely moving biological flocculators. Straight chemical precipitation in either the primary or final clarifiers is recommended for trickling filters in preference to direct dosing across the filter media.

EXCESS SLUDGE PRODUCTION WITH MINERAL ADDITION

Excepting any improvement effected in secondary clarification, the additional sludge produced via mineral addition can be attributed to the precipitation of mineral-phosphate complexes. The quantity of excess sludge so generated is a legitimate concern of treatment plant designers.

Two recently conducted studies provide information on this subject. The first was conducted at the joint FWQA-Prince William County, Virginia pilot plant in Manassas, Virginia using only aluminum salts (7). Excess sludge production was found to be a straight line function up to the stoichiometric requirement of aluminum ion. Stoichiometric requirement is defined as that amount of aluminum ion which yields a 1/1 mole ratio of Al/P. Above the stoichiometric dose, excess sludge production tailed off sharply. Figure 20 shows that at Manassas 4 mg/l of excess sludge were produced for every mg/l of aluminum ion added up to the stoichiometric dose of 13.5 mg/l of Al. However, it was also found that improvement in sludge volume index and sludge dewaterability due to the aluminum compensated for increased sludge mass to the extent that net volume of waste sludge remained the same and dewatering capacity did not have to be enlarged.

Both aluminum and iron salts were fed to a small package plant in New Berlin, Wisconsin during a programmed study in 1968 (8). Additional solids production ranged from 7-8 pounds per pound of phosphorus removed throughout although one would intuitively expect more excess sludge with iron due to its higher atomic weight. The increase in the total weight of solids wasted was approximately 100%. However, as at Manassas, a considerable improvement in mixed liquor settling rate, settled volume, and drainability was noted. Volumetric waste sludge requirements were increased by only 35%.

COST CONSIDERATIONS FOR MINERAL ADDITION

Under present pricing structures, alum is less costly than aluminate. Figure 21 indicates that for equivalent aluminum ion content, powdered alum is about 15% cheaper than powdered aluminate and that liquid alum is approximately 35% cheaper than liquid aluminate. The prices shown here are bulk delivered prices to the Greater Cincinnati area from the nearest production site for each chemical.

The alum pricing structure is very stable owing to an extensive existing market in the water treatment and paper industries. On the other hand, aluminate pricing is much more elastic and the opening of a major new market in wastewater treatment could quite conceivably lower the bulk purchase rate significantly. Widespread utilization of the mineral addition technique for phosphorus removal would almost certainly diminish the current price differential in the two compounds.

Reduced to a unit cost basis, the chemical cost for 80-90% phosphorus removal using aluminum salts will range from about $30-50 per million gallons treated for an influent phosphorus concentration of 10 mg/l as P.

Figure 22 lists the various options for the two levels of desired removal. Liquid alum is the least expensive; liquid aluminate is the most expensive. An alum + lime option is included for those plants which have low alkaline wastewater but may choose to use lime doses to the aerator to replenish alkalinity depleted by the alum rather than switch to the basic but more costly aluminate. As shown in Figure 23, ignoring the increased total dissolved solids added thereby, alum + lime is still somewhat cheaper than aluminate alone using the cheapest present day source of both aluminum products.

Chemical costs for iron compounds varies considerably with the location and the availability of waste pickle liquor. Commercial grade iron compounds will cost approximately the same or slightly less than alum for equivalent phosphorus removal. Waste pickle liquor, though, if available, will bring the chemical cost down to a fraction of the cost using high grade precipitants. The University of Texas research grant project is a good example (5). The Texas City, Texas Plant No. 2 utilized for the study has a normal phosphorus load of about 10 mg/1 as P. To obtain 90% phosphorus removals, a mole ratio of Fe/P of 1.5/1 was required. Using waste ferrous sulfate pickle liquor ($FeSO_4 \cdot 7H_2O$) at the regular delivered price of $0.095 per pound of available Fe, the chemical cost would have been $20 per million gallons treated (2¢ per thousand gallons). However, the waste pickle liquor was made available to the project for the cost of trucking only at $0.015 per pound of available Fe. The resulting chemical cost was $3.2 per million gallons treated (0.32¢ per thousand gallons), or about 1/6 the cost if pickle liquor had been purchased at the regular price.

A relatively modest investment will purchase the necessary capital equipment for mineral addition. The requirements are storage tanks, feed pumps and plumbing, and for accurate chemical proportioning to incoming phosphorus a controlled dosing system. One variation of a controlled dosing system is shown in Figure 24. Hydraulic flow and incoming phosphate are continuously monitored. A positive displacement controlled volume pump is utilized to meter in the liquid chemical. The drive speed of the pump is automatically controlled as a coarse adjustment by an electrical signal from the flow recorder. A signal from the continuous phosphate analyzer is used to automatically vary piston stroke length as the fine adjustment. In this manner, the ratio of chemical fed to influent phosphorus is always maintained at a predetermined level. The system described above is a dual-automatic open loop system. The critical piece of hardware is the continuous on-line phosphate analyzer. Individual investigators have modified laboratory analyzers to serve in an on-line capacity but, to date, no company is marketing an on-line model.

The total cost of phosphorus removal via mineral addition including chemicals, O & M, and debt retirement is estimated in Figure 25. In this example, liquid alum is costed in at the dose required to remove 90% of a phosphorus load of 10 mg/l as P. It is readily apparent that chemicals constitute the major cost element, particularly for larger plants. Incremental costs for phosphorus removal over and above the total annual operating costs for normal operation of an activated sludge plant are estimated at roughly 25-50%, depending on plant size. Clearly, if we are going to consider phosphorus a major pollutant of the same magnitude as organic carbon and suspended solids, we have to be willing to pay a price for its removal which approaches the unit cost of either organic carbon or suspended solids removal.

FWQA GRANT AND CONTRACT PROJECTS USING MINERAL ADDITION

Itemized below is a list of current FWQA projects using mineral addition to activated sludge systems to remove phosphorus. Several of the projects are in the design or construction phase and are not yet generating data.

1. Penn State University - 2 mgd (alum)
2. University of Texas (Texas City) - 0.75 mgd (waste pickle liquor)
3. University of Notre Dame - 50,000 gpd (aluminate)
4. Manassas, Virginia - 1 mgd (aluminate)
5. New York State Environmental Conservation Department (West Coxsackie, N. Y.) - 0.1 mgd (aluminate)
6. Detroit, Michigan - 7,000 gpd (waste pickle liquor)
7. Xenia, Ohio (Little Miami) - 1 mgd (aluminate)
8. Seattle, Washington (METRO) - 20 mgd (aluminum salts)
9. New York City - 20 mgd (alum)
10. Milwaukee, Wisconsin - 80 mgd (iron salts)

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8. Schmit, F. L. and Ewing, L., "Phosphate Removal Systems for Small Activated Sludge Plants," Pennsylvania Water Pollution Control Assn., (August 1970).

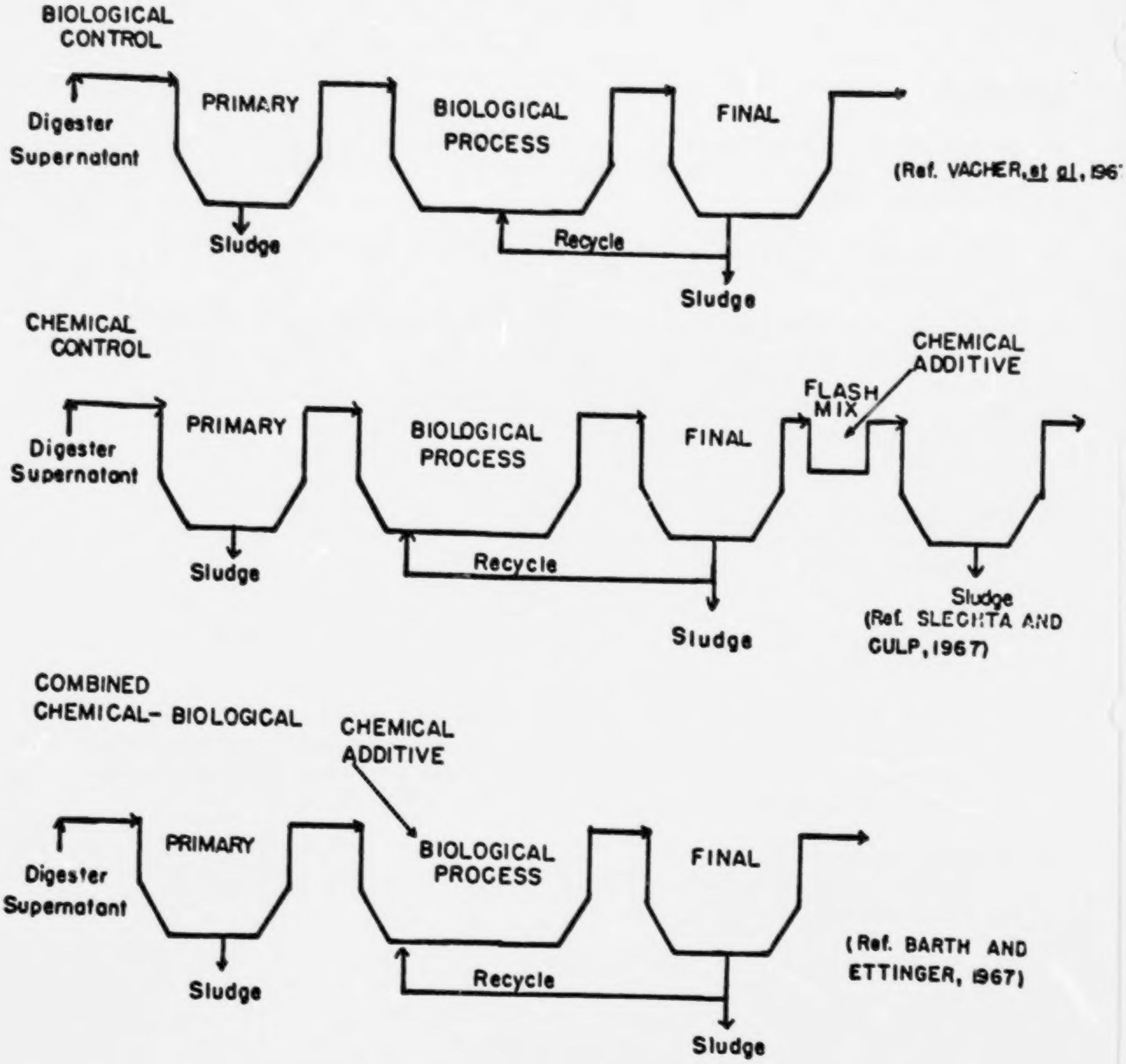

BIOLOGICAL CONTROL
Digester Supernatant
PRIMARY
BIOLOGICAL PROCESS
FINAL
Sludge
Recycle
Sludge
(Ref. VACHER, et al, 196
CHEMICAL CONTROL
Digester Supernatant
PRIMARY
BIOLOGICAL PROCESS
FINAL
FLASH MIX
CHEMICAL ADDITIVE
Sludge
Recycle
Sludge
Sludge
(Ref. SLECHTA AND CULP, 1967)
COMBINED CHEMICAL- BIOLOGICAL
CHEMICAL ADDITIVE
Digester Supernatant
PRIMARY
BIOLOGICAL PROCESS
FINAL
Sludge
Recycle
Sludge
(Ref. BARTH AND ETTINGER, 1967)

FIGURE 1. PHOSPHORUS CONTROL BASIC SYSTEMS

FIGURE 2

CHOICE OF CHEMICALS FOR PHOSPHORUS REMOVAL

PRIMARY

1. FERROUS CHLORIDE + BASE + POLYMER
2. FERRIC CHLORIDE + POLYMER
3. FERROUS SULFATE + BASE + POLYMER
4. FERRIC SULFATE + POLYMER
5. WASTE PICKLE LIQUOR + BASE + POLYMER
6. SODIUM ALUMINATE + POLYMER
7. ALUMINUM SULFATE (ALUM) + POLYMER
8. LIME (SINGLE STAGE)
9. LIME (TWO STAGE)

SECONDARY

1. SODIUM ALUMINATE
2. ALUM
3. ABOVE IRON COMPOUNDS

TERTIARY

1. LIME (SINGLE STAGE)
2. LIME (TWO STAGE)
3. ALUM
4. ACTIVATED ALUMINA COLUMNS

FIGURE 4.

ADVANTAGES OF
COMBINED BIOLOGICAL AND CHEMICAL PROCESSES

Add Mineral Supplement Directly to Aerator

---

1. Dual Use of Aerator
    - a) Mixing by Diffusion Apparatus
    - b) Residence Time for Reactions
2. Uses Existing Final Settler
3. Large Surface Area of Floc Available
    - a) Biological - Chemical Interaction
    - b) Reduces Chemical Requirement
4. Improves Mixed Liquor Settling
5. Reduces Effective Organic Loading
6. Reduces Supernatant Phosphorus Recycle
    - a) Phosphorus not Released during Digestion
    - b) Phosphorus Removed from Plant in Digested Sludge

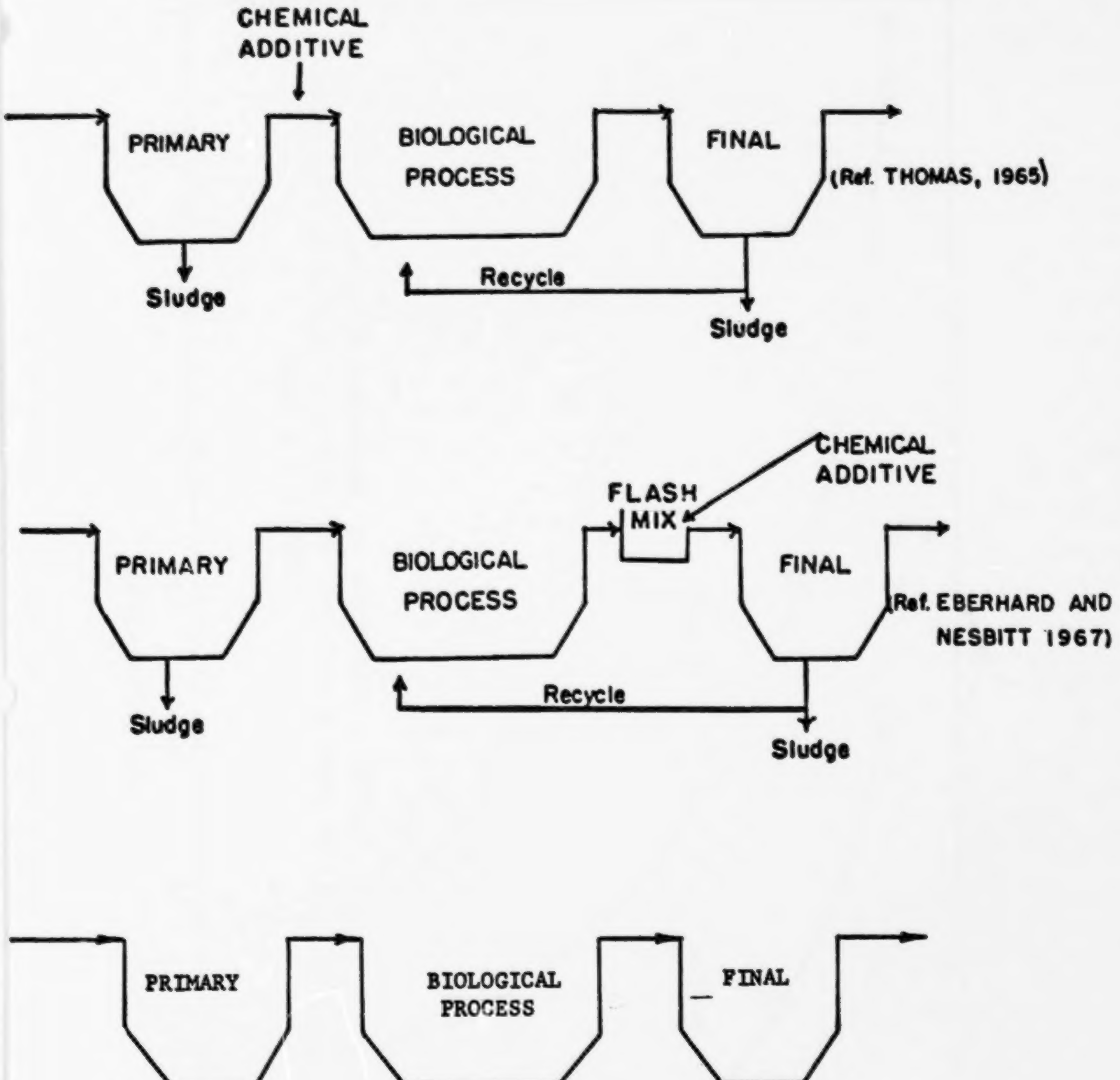

CHEMICAL ADDITIVE
PRIMARY
BIOLOGICAL PROCESS
FINAL
(Ref. THOMAS, 1965)
Sludge
Recycle
Sludge
CHEMICAL ADDITIVE
FLASH MIX
PRIMARY
BIOLOGICAL PROCESS
FINAL
(Ref. EBERHARD AND NESBITT 1967)
Sludge
Recycle
Sludge
PRIMARY
BIOLOGICAL PROCESS
FINAL
Recycle
Sludge
Sludge
CHEMICAL ADDITIVE

FIGURE 3. COMBINED CHEMICAL- BIOLOGICAL, VARIATIONS OF BASIC SYSTEMS

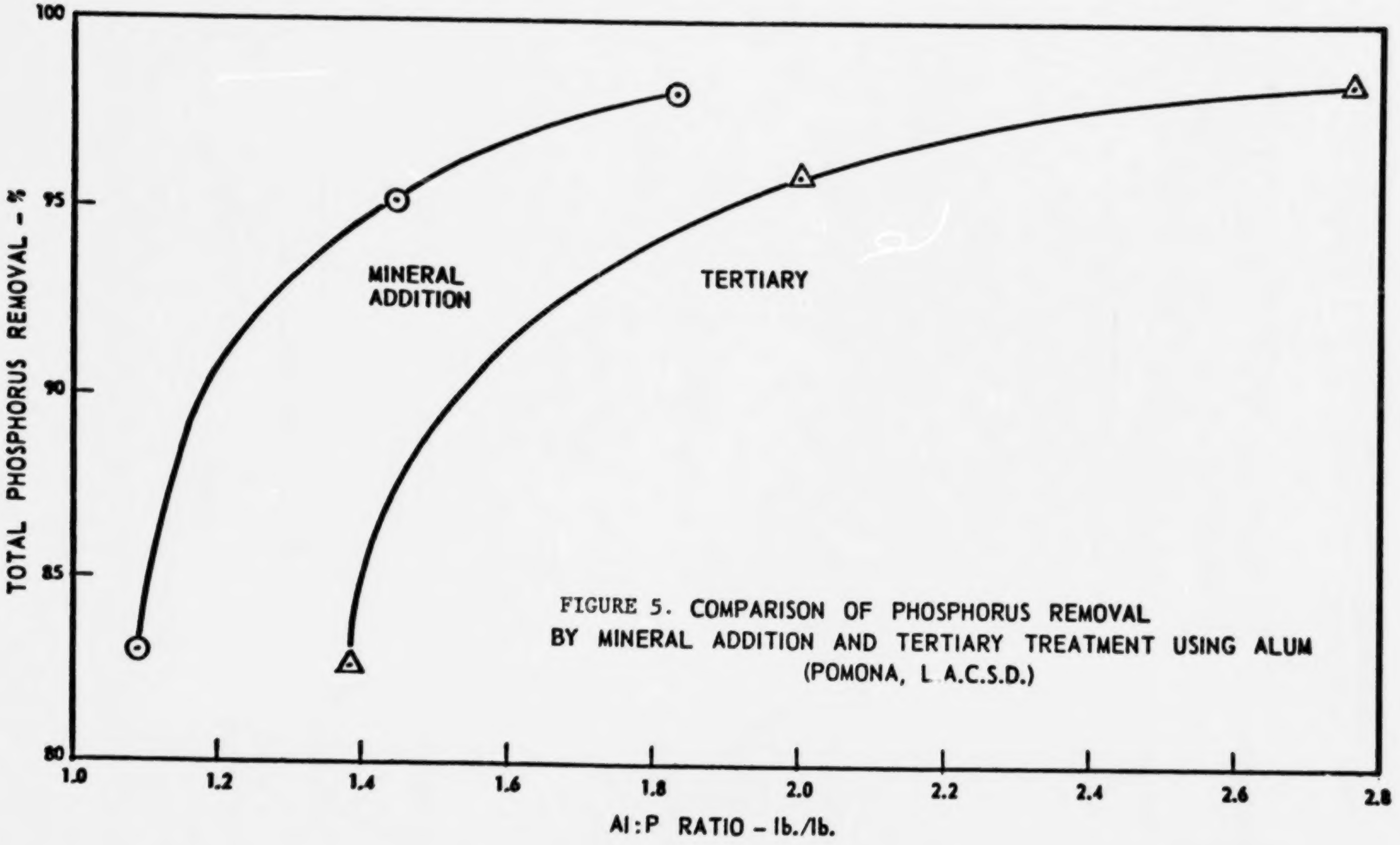


FIGURE 5. COMPARISON OF PHOSPHORUS REMOVAL BY MINERAL ADDITION AND TERTIARY TREATMENT USING ALUM (POMONA, L.A.C.S.D.)

FIGURE 6.

EFFECT OF ALUMINUM ON SLUDGES AND SUPERNATANT

1. HIGH-RATE MIXED LIQUOR CHARACTERISTICS

	MLSS mg/l	MLVSS mg/l	% VOLATILE	SDI	ORGANIC LOAD lb COD/day/ lb MLVSS
BEFORE Al^{+++}	1,450	900	62	0.44	2.6
AFTER Al^{+++}	3,385	1,740	51	1.12	1.6

2. PERCENT PHOSPHORUS IN SLUDGES - TOTAL SOLIDS BASIS

	PRIMARY	WASTE ACTIVATED	MIXED PRIMARY PLUS W. ACTIVATED	DIGESTED
BEFORE Al^{+++}	1.1	2.1	1.6	2.2
AFTER Al^{+++}	1.1	6.2	4.5	5.2

3. TOTAL PHOSPHORUS IN SUPERNATANT - mg/l

BEFORE Al^{+++}	170
AFTER Al^{+++}	60

FIGURE 7.

CHARACTERISTICS OF ALUMINUM SALTS

	(ALUM) ALUMINUM SULFATE		SODIUM ALUMINATE	
CHEMICAL FORMULA	$Al_2(SO_4)_3$		$Na_2Al_2O_4$ ($NaAl(OH)_4$ DISSOLVED)	
pH of 1% SOLUTION OF DRY CHEMICAL	3.5		11.9	
	DRY FORM	LIQUID FORM	DRY FORM	LIQUID FORM
% STATED CHEMICAL WT. BASIS	57	28	74	46
% Al - WT. BASIS	9.0	4.4	24.4	15.1

FIGURE 8.

SIDE EFFECTS OF MINERAL ADDITION

COMPOUND	EXTRANEOUS ION ADDED	LB EXTRANEOUS ION ADDED PER LB METAL ION *
ALUM	SO_4^{--}	5.35
SODIUM ALUMINATE	Na^+	0.85
FERROUS CHLORIDE	Cl^-	1.26
FERRIC CHLORIDE	Cl^-	1.91
FERROUS SULFATE	SO_4^{--}	1.72
FERRIC SULFATE	SO_4^{--}	2.58

*1 lb of Al ion is equivalent to 2.07 lb of Fe ion for precipitation of phosphorus.

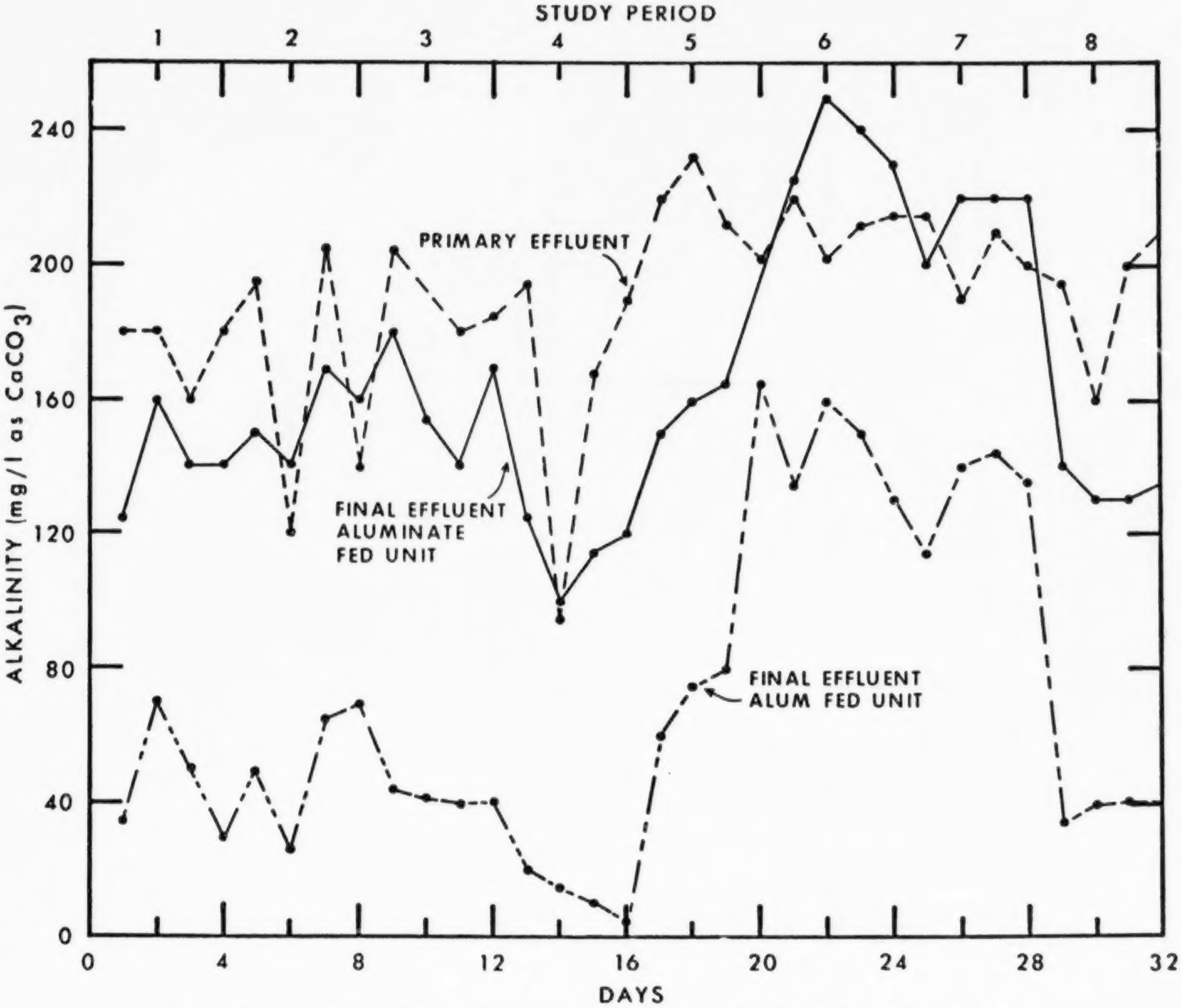


FIGURE 9. EFFECT OF MINERAL ADDITION ON ALKALINITY OF SYSTEMS

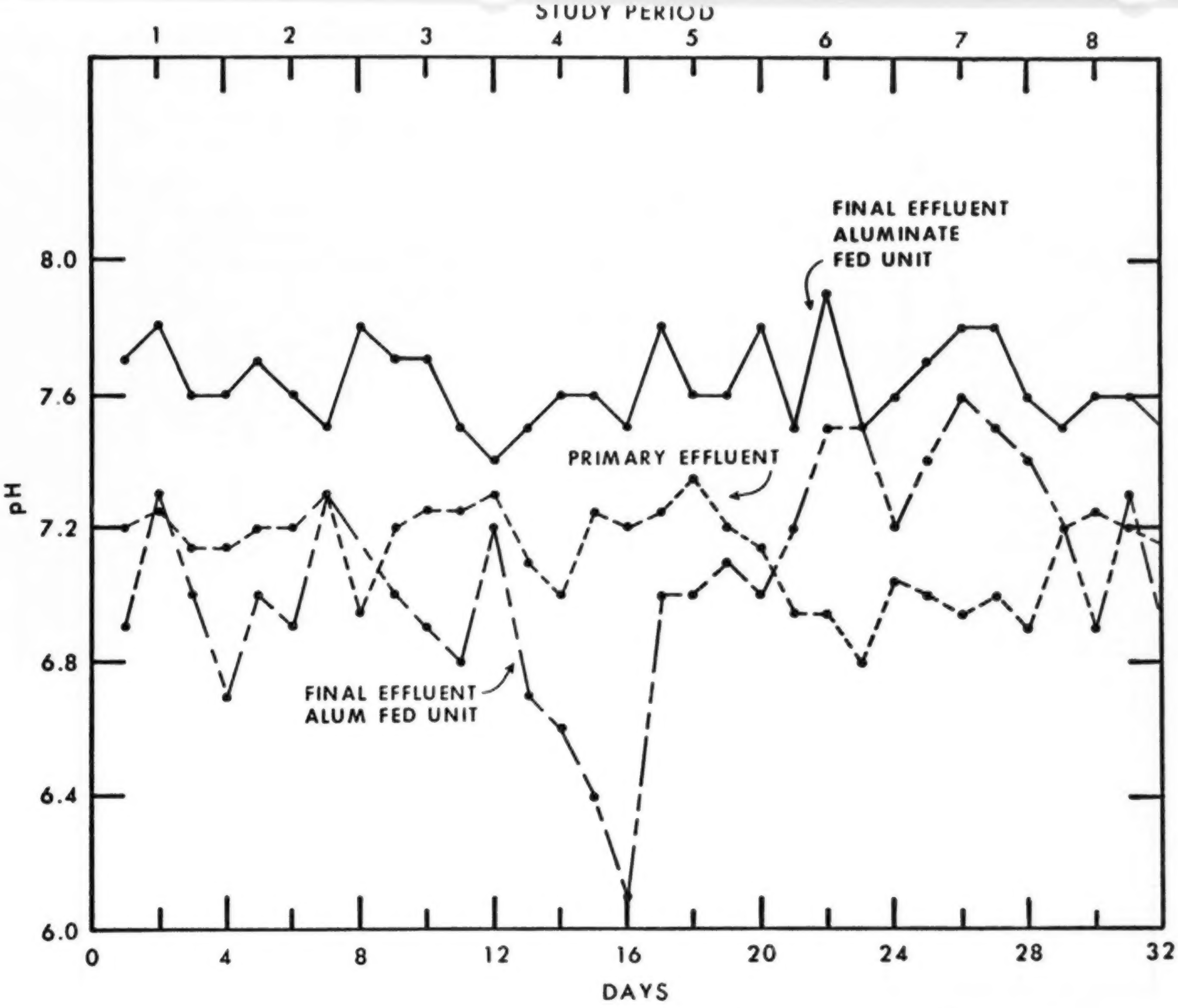


FIGURE 10 EFFECT OF MINERAL ADDITION ON pH OF SYSTEMS

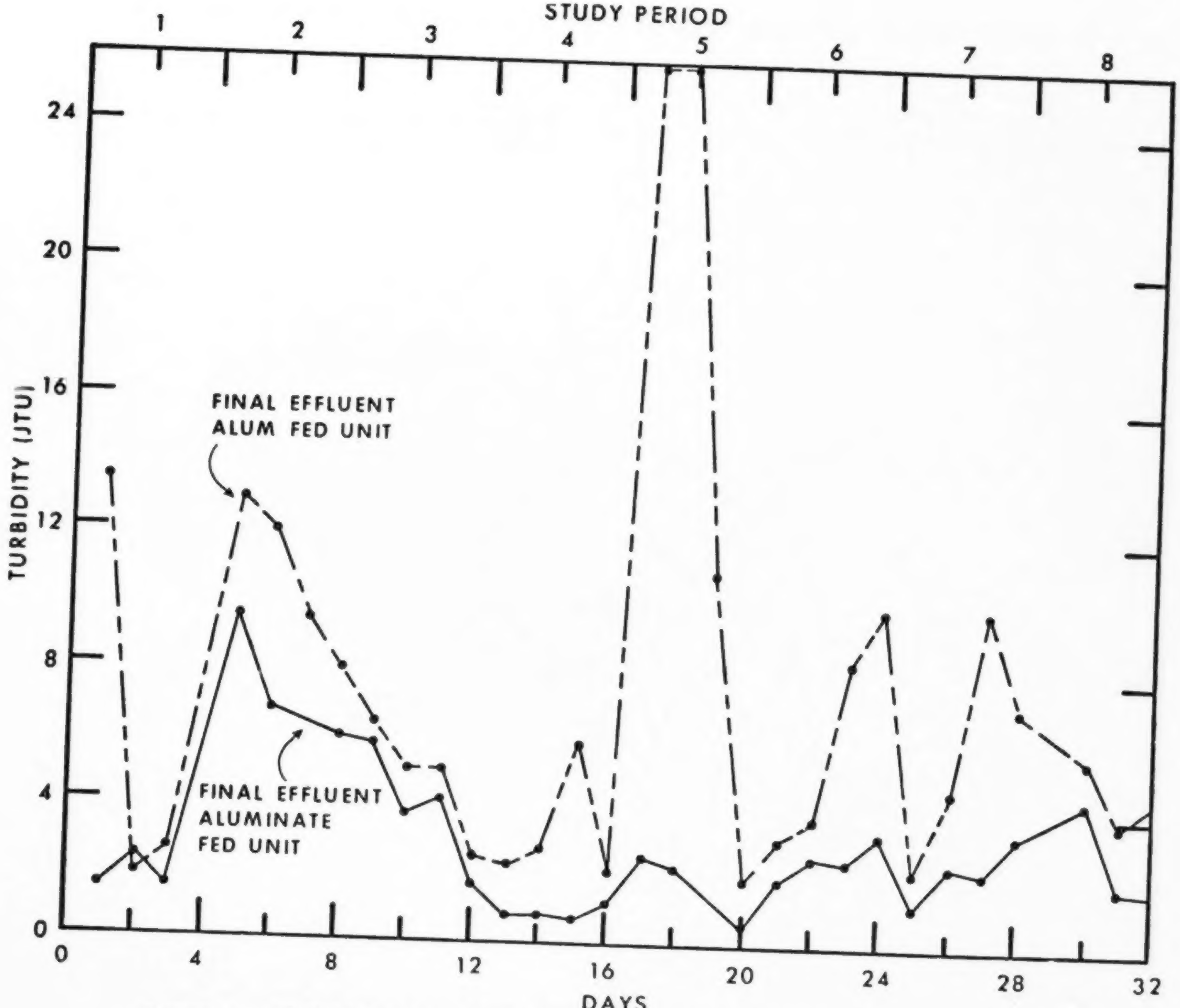


FIGURE 11. COMPARISON OF EFFLUENT TURBIDITY

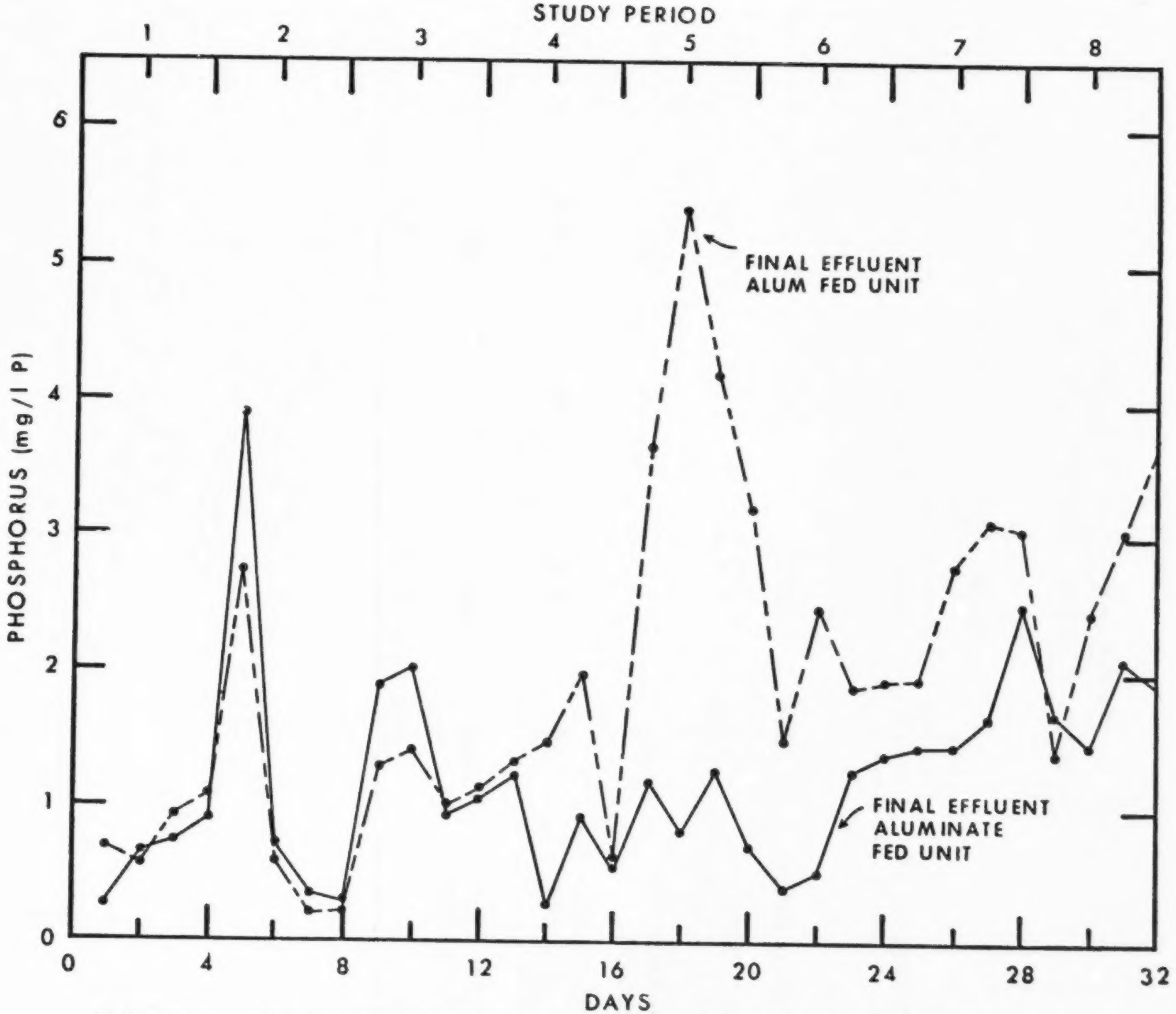


FIGURE 12. COMPARISON OF EFFLUENT PHOSPHORUS

FIGURE 13.

COMPARISON OF ALUM DOSED AND ALUMINATE DOSED ACTIVATED SLUDGE UNITS

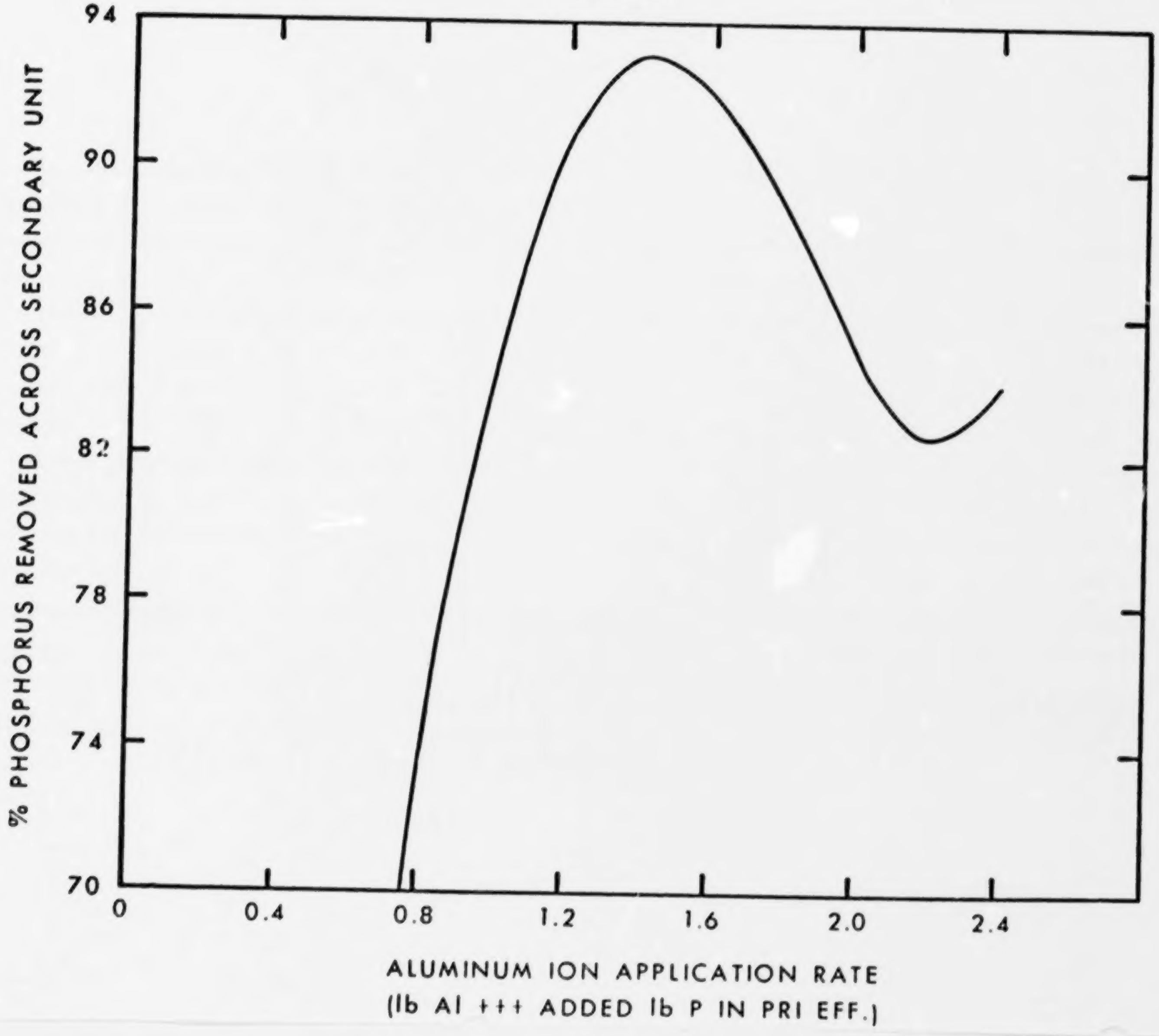
CONSTITUENT	RAW SEWAGE mg/l	FINAL EFF - ALUMINATE DOSED UNIT		FINAL EFF - ALUM DOSED UNIT	
		mg/l	% REMOVAL	mg/l	% REMOVAL
COD	310	23	93	35	89
TOTAL PHOS.	11.6	1.2	90	2.0	83
NITROGEN (ALL FORMS)	23.0	9.4	59	10.7	53
SUSPENDED SOLIDS	157	7	96	21	87
TURBIDITY - PRI EFF - (JTU)	46	3.1	93	7.4	85
ALKALINITY (as $CaCO_3$)	182	165		67	
pH	7.2	7.6		6.9	

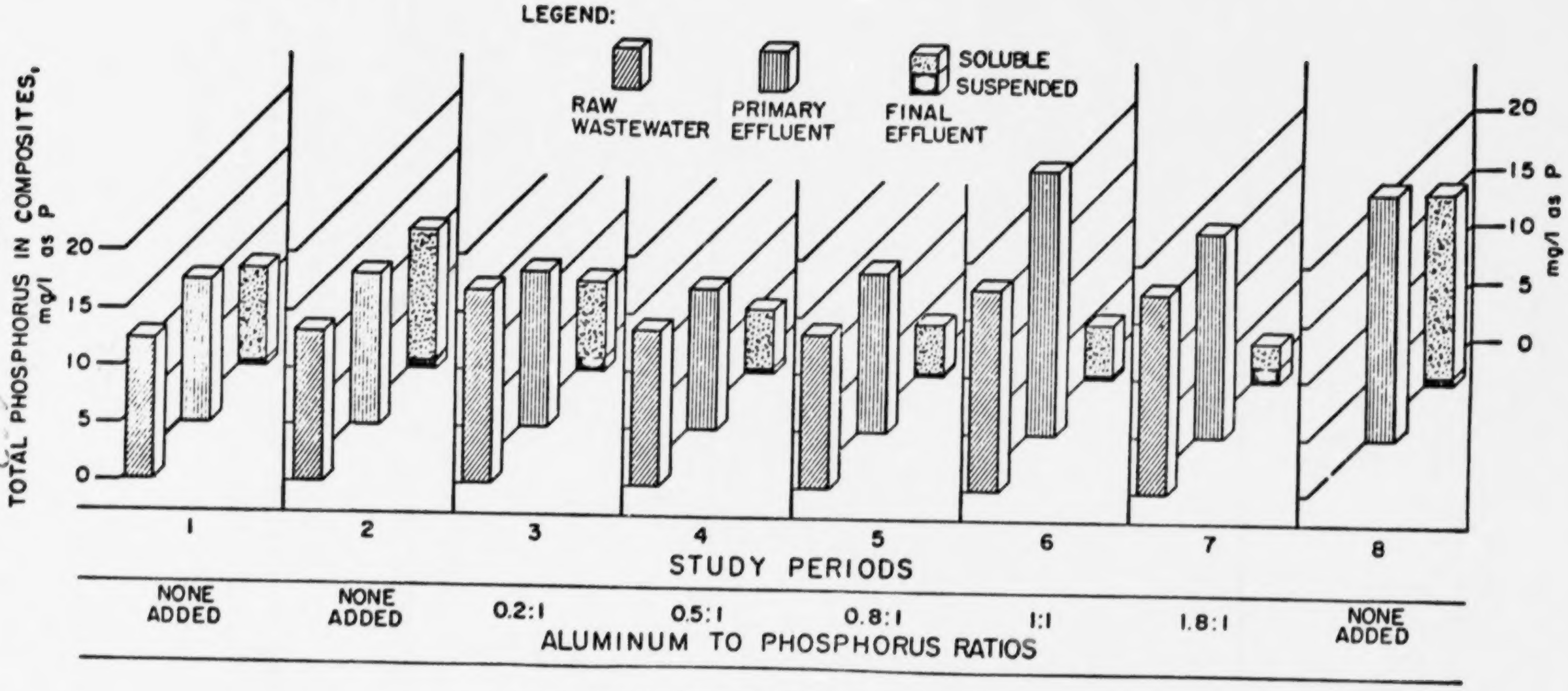
FIGURE 14.

POMONA DATA, AVERAGE VALUES

1968 DATE	CHEMICAL ADDED	AL:P RATIO WT./WT.	Phosphorus, mg/l			Percent Removal	
			PRIMARY EFFLUENT	FINAL EFFLUENT (UNFILTERED)	FINAL EFFLUENT (FILTERED)	UNFILTERED	FILTERED
2/12 - 2/26	ALUMINATE	1:1	13.2	2.6	2.1	80	83
2/26 - 3/11	ALUMINATE	1.5:1	13.2	2.3	0.9	83	94
3/18 - 4/1	ALUM	1.1:1	12.8	2.7	2.2	79	83
4/1 - 4/15	ALUM	1.4:1	13.5	1.4	0.7	90	95
4/15 - 4/29	ALUM	1.8:1	13.8	1.5	0.3	89	98
5/13 - 5/27	ALUMINATE	1.5:1	13.2	2.4	1.4	82	90

FIGURE 15.


94
90
86
82
78
74
70
0
0.4
0.8
1.2
1.6
2.0
2.4
% PHOSPHORUS REMOVED ACROSS SECONDARY UNIT
ALUMINUM ION APPLICATION RATE
(lb Al +++ ADDED lb P IN PRI EFF.)



LITTLE MIAMI TREATMENT PLANT (1 mgd)

FIGURE 16. AVERAGE PHOSPHORUS VALVES FOR 5-DAY PERIODS

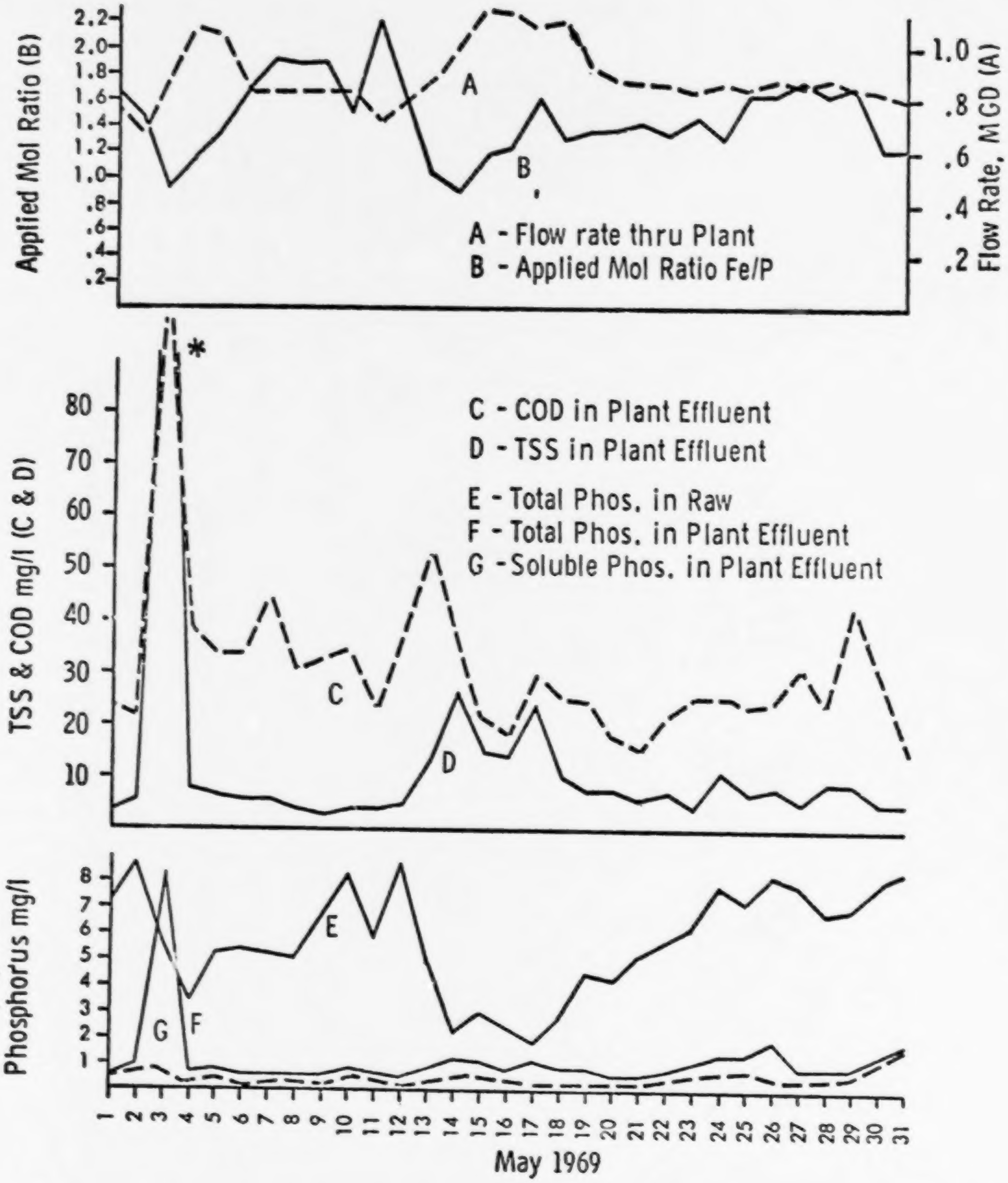


FIGURE 17. _ Ferrous Iron injection to raw influent, Texas City Plant No. 2 May 1969 *(on May 3 inadvertant acceptance of 1.5-1.6 MGD rate for 2-3 hours, raised effluent COD, suspended solids and phosphorus for the day to 108, 141, and 8.3 mg/l respectively.)

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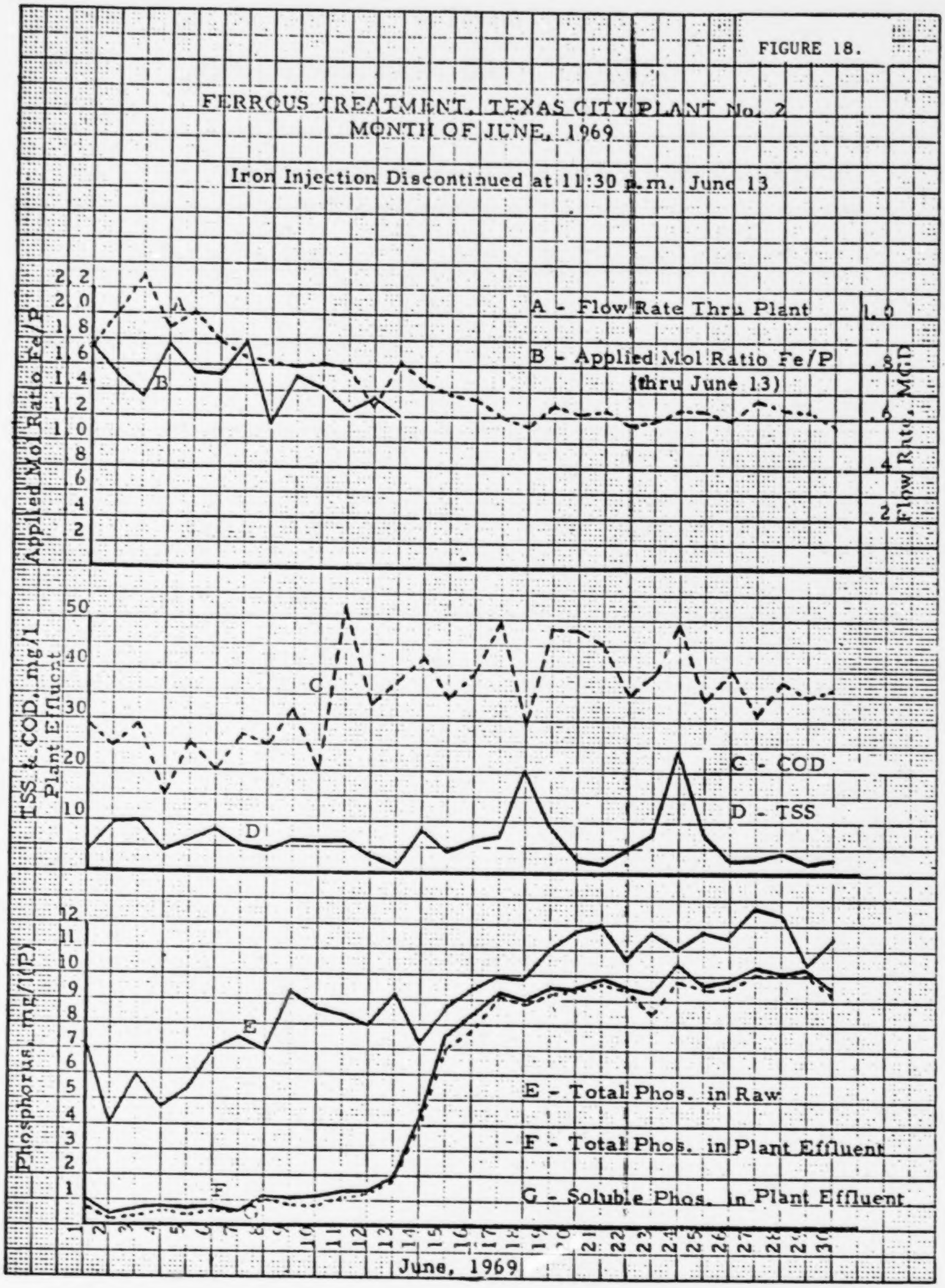
FIGURE 18.



FERROUS TREATMENT, TEXAS CITY PLANT No. 2
MONTH OF JUNE, 1969

Iron Injection Discontinued at 11:30 p.m. June 13

FIGURE 19.

FAIRBORN, OHIO TRICKLING FILTER
(Percent Removals)

Unit efficiency - Primary effluent to final effluent

	Total Phosphorus	COD	S. S.	pH Mode	Al : P
Dosed Filter	63	74	51	7.9	1 : 1
Control Filter	13	75	51	7.8	--

Overall efficiency - Raw waste water to final effluent

	Total Phosphorus	COD	S. S.	pH Mode	Al : P
Dosed Filter	64	83	84	7.9	1 : 1
Control Filter	17	83	85	7.8	--
Entire Plant	37	84	92	7.6	0.3 : 1

FIGURE 20. ADDITION SLUDGE PRODUCED WITH MINERAL ADDITION AT MANASSAS, VIRGINIA

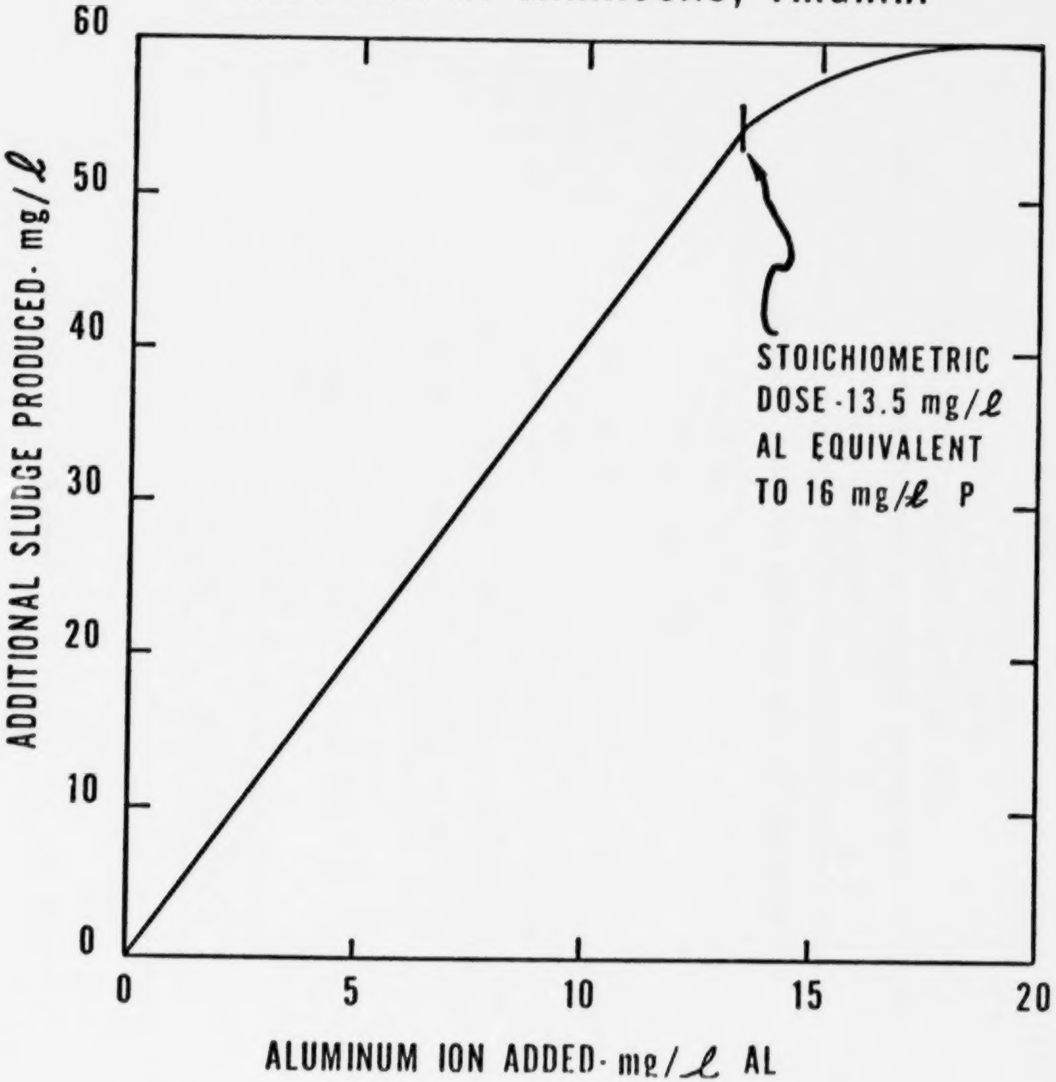


FIGURE 21

BULK DELIVERED CHEMICAL PRICE STRUCTURE FOR PROJECTING PHOSPHORUS REMOVAL COSTS IN THE MIDWEST

(JANUARY, 1971)

ALUMINUM SULFATE (ASSUMED 100-MILE DELIVERY)

1. LIQUID - $26/ton as PRODUCT
   $600/ton as Al^{+++}
2. DRY - $65/ton as PRODUCT
   $720/ton as Al^{+++}

SODIUM ALUMINATE (ASSUMED 300-MILE DELIVERY)

1. LIQUID - $115/ton as PRODUCT
   $810/ton as Al^{+++}
2. DRY - $182/ton as PRODUCT
   $750/ton as Al^{+++}

QUICKLIME (CaO) - $24/ton

FIGURE 22

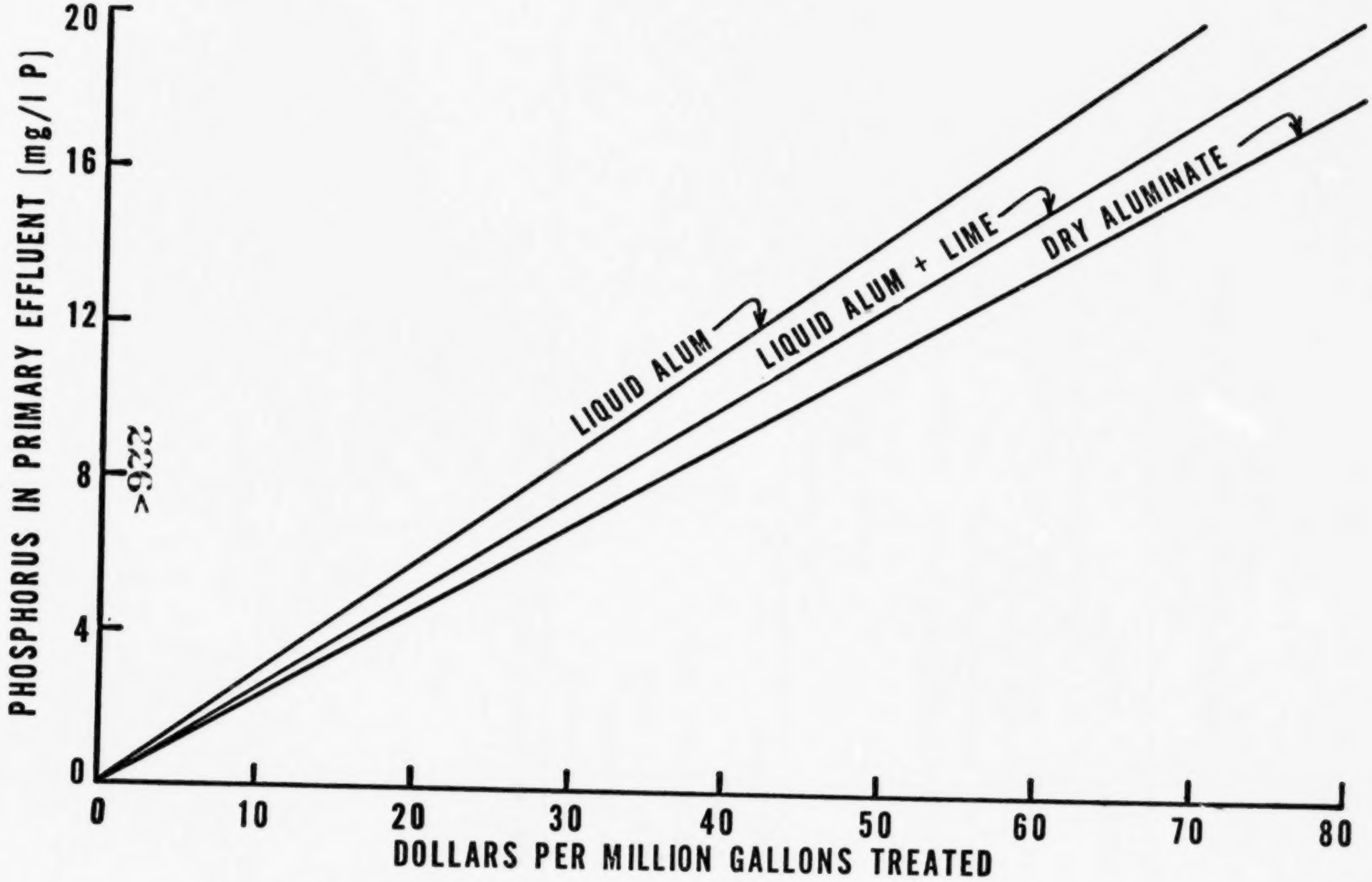
ESTIMATED
CHEMICAL COSTS OF REMOVING PHOSPHORUS
BY MINERAL ADDITION

BASIS:

1. PRIMARY EFFLUENT PHOSPHORUS = 10 mg/l as P.
2. EVERY 5 mg/l ALUM ADDED (as Al^{+++}) USES UP 34 mg/l ALKALINITY (as $CaCO_3$).
3. FOR 80% PHOSPHORUS REMOVAL, Al/P = 1.2/1 BY WEIGHT.
4. FOR 90% PHOSPHORUS REMOVAL, Al/P = 1.4/1 BY WEIGHT.

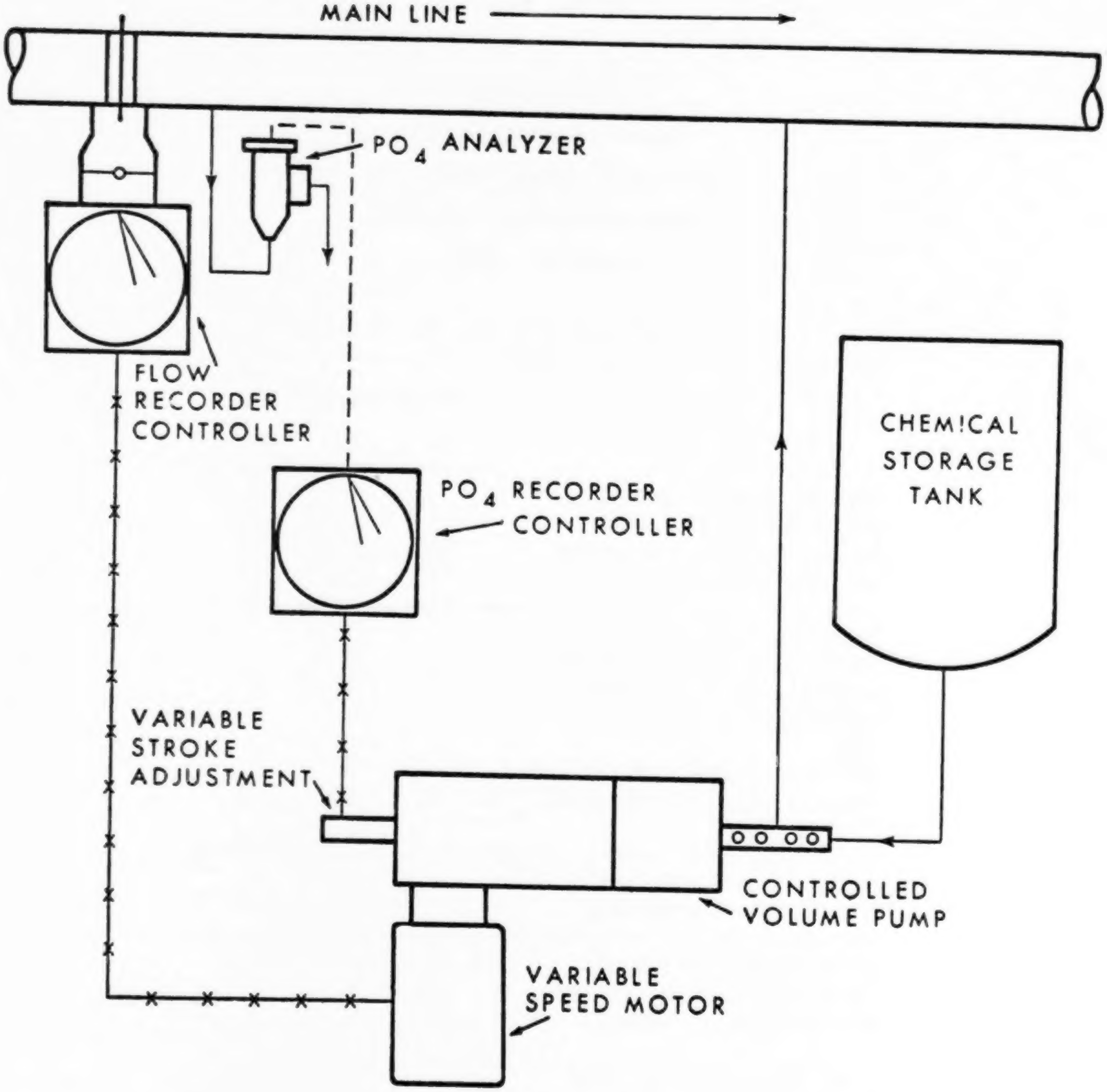
	ALL COSTS IN DOLLARS PER MILLION GALLONS TREATED	
	80% REMOVAL	90% REMOVAL
LIQUID ALUM	30	35
DRY ALUM	36	42
LIQUID ALUM + LIME*	35	40
DRY ALUM + LIME*	41	45
DRY ALUMINATE	38	44
LIQUID ALUMINATE	41	47

*FOR WASTEWATERS WITH LOW ALKALINITY



CHEMICAL COST FOR 90% PHOSPHORUS REMOVAL

FIGURE 23


MAIN LINE
PO_4 ANALYZER
FLOW RECORDER CONTROLLER
CHEMICAL STORAGE TANK
PO_4 RECORDER CONTROLLER
VARIABLE STROKE ADJUSTMENT
CONTROLLED VOLUME PUMP
VARIABLE SPEED MOTOR

DUAL AUTOMATIC OPEN LOOP
CHEMICAL FEED SYSTEM FOR CONTROLLED DOSING
FIGURE 24.

FIGURE 25

ESTIMATED
TOTAL TREATMENT COSTS
ACTIVATED SLUDGE TREATMENT
PLUS PHOSPHORUS REMOVAL
(JANUARY 1970)

COSTS IN DOLLARS PER MIL GAL TREATED

	PLANT SIZE - MGD		
	1	10	100
CURRENT ANNUAL OPERATING COSTS FOR CARBON AND SOLIDS REMOVAL INCLUDING DEBT RETIREMENT	233	135	82
CHEMICAL COST FOR 90% PHOSPHORUS REMOVAL USING LIQUID ALUM - INFLUENT PHOSPHORUS = 10 mg/1 P	35	35	35
ADDITIONAL DEBT RETIREMENT AND O & M COSTS FOR PHOS. REMOVAL	25	10	5
TOTAL INCREMENTAL COSTS FOR PHOSPHORUS REMOVAL	60	45	40
TOTAL ANNUAL OPERATING COSTS FOR CARBON, SOLIDS AND PHOSPHORUS REMOVAL	293	180	122
% INCREASE IN TREATMENT COSTS FOR PHOSPHORUS REMOVAL	26	33	49

BACTERIAL AND VIRAL CONTROL AND WATER REUSE
(for domestic purposes)

by

Harold W. Wolf, Dr.P.H.*

Introduction

The problems that we face in any consideration of reusing treated wastewaters for domestic purposes, even if they be limited to the microbiological, are really quite demanding. The reason for this difficulty is basically one of philosophy but a philosophy that carries over to the technical area. It is not generally realized, but the U.S. Public Health Service Drinking Water Standards, upon which we must ultimately evaluate qualities of treated wastewaters, is, to a great extent a standard of expedience. It does not exclude entirely the possibility of acquiring an intestinal infection. It is attainable by the economic development of available water supplies, their disinfection, and if need be treatment in purification works by economically feasible methods. It is not a standard of perfection. Fair & Geyer[1] note that the application of the Standard to areas of high endemicity of diseases that are transmittible through drinking water is challengeable.

In any consideration of drinking treated wastewaters we cannot accomodate a philosophy of expedience. We must be dead certain that no one is going to become ill by virtue of drinking this type of water. This added degree of certainty is really what makes the microbiological problem so difficult. Disinfection is a rate process and it is a complicated rate process dependent upon the physico chemistry of the disinfectant, the cyto-chemical nature and physical state of the pathogens, the reaction of the two, and quantitative effects of factors in the reaction medium such as temperature, pH, electrolytes, and interfering substances. Coupling the absolute certainty of non-infectivity with the wide

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* Director, Dallas Water Reclamation Research Center, Texas A&M University, Dallas, Texas

variation that occurs in the quality of wastewaters plus the even broader variations in numbers and types of pathogenic organisms, one must admit that the problems are, indeed, formidable.

Meeting Drinking Water Standards

We have all repeatedly seen references to articles proclaiming that sewage has been treated to such a high degree that it meets Drinking Water Standards. Such statements are not accurate. What is meant is that they meet some of the requirements of Drinking Water Standards. There are two paragraphs (the first two definitions, in fact) in the Drinking Water Standards that are very much a part of those Standards. The requirements of these paragraphs must be met in order to render the judgment that a specific product water "meets Drinking Water Standards". The first refers to "Adequate protection by natural means" which "involves one or more of the following processes of nature that produces water consistently meeting the requirements of these Standards: dilution, storage, sedimentation, sunlight, aeration, and the associated physical and biological processes which tend to accomplish natural purification in surface waters, the natural purification of water by infiltration through soil and percolation through underlying material and storage below the ground water table".

The second paragraph concerns "Adequate protection by treatment " which means "any one or any combination of the control processes of coagulation, sedimentation, absorption, filtration, disinfection, or other processes which produce a water consistently meeting the requirements of these Standards. This protection also includes processes which are appropriate to the source of supply; works which are of adequate capacity to meet maximum demands without creating health hazards, and which are located, designed and constructed to eliminate or prevent pollution; and conscientious operation by well-trained and competent personnel whose qualifications are commensurate with the responsibilities of the position and acceptable to the Reporting Agency and the Certifying Authority".

Since "adequate protection by treatment" has not been defined by the Public Health Service, which is the Certifying Authority referred to, there are no such waters that "meet Drinking Water Standards".

It has already been mentioned that Fair and Geyer state that the application of the coliform standard is questionable in areas of high endemicity of waterborne diseases. It can therefore be stated that the Standard is also similarly questionable if it is to apply to a source water when that source water is sewage.

Viruses

The Committee on Environmental Quality Management of the Sanitary Engineering Division of the American Society of Civil Engineers has recently published a paper "Engineering Evaluation of Virus Hazard in Water"[2]. The paper quotes the following analysis of the significance of low concentrations of viruses by Hillel Shuval of Israel:

"The findings of Dr. Lund, confirmed by the research in South Africa, that sewage treatment plants reduced the viral content of sewage by one or possibly by two logs, added to the estimate of Dr. Coin that one infectious dose of virus was present in 250 liters of Paris drinking water, provoke speculation. If the viral concentration in a sewage is 10 $TCID_{50}$ per ml and this amount is reduced as much as two logs in a sewage treatment plant, another three logs by dilution in a river, and two or three more logs in a water treatment plant, a total of seven or eight log reduction in viral concentration results. This leaves a water containing perhaps one $TCID_{50}$ of virus per 1000 liters. I suggest that this is not an unimportant amount. It is true that one person does not drink 1000 or even 250 liters of water per day, but one in 250 or one in 1000 or even one in 10,000 persons ingests one $TCID_{50}$ of virus daily and one $TCID_{50}$ is an infectious dose for man, as Dr. Plotkin suggested. Even after the viral concentration has been reduced 99.9999 percent, we have a level of ingestion of virus that could be responsible for a proportion of the endemic viral diseases in communities consuming highly treated waters."

It is therefore quite obvious that in any consideration of using highly treated wastewaters as a domestic supply, the treatment processes must be extremely efficient in destroying viruses.

Microbiological Control

The same A.S.C.E. publication then asks the question "Can (the) coliform index be reasonably applied to (an) evaluation of enteric viruses in water?" In response to this question, the committee quotes Clarke[3] extensively as follows:

"In order to answer this question, the following factors must be considered:

"1. The relative numbers of indicator bacteria and viruses in water and sewage must be known.

"2. The relative effectiveness of water and sewage treatment processes on bacteria and viruses must be determined.

"3. The relative survival time of the two groups of microorganisms must be evaluated.

"4. Is there epidemiological evidence that water which does or does not meet bacteriological standards can be responsible for infection and/or disease in humans?

"The relative numbers of coliform bacteria and viruses in water and sewage have been discussed by Clarke et al.[4] These authors have calculated that the average enteric virus density in domestic sewage is probably about 500 virus units per 100 ml .... coliform densities in domestic sewage average 46 x 10^6 per 100 ml... thus, coliform-virus ratios in sewage are about 92,000 to 1.... It is apparent that coliform organisms far outnumber enteric viruses in sewage... and therefore are a far better indicator of pollution than are enteric viruses.

"The next point to be considered is the relative effectiveness of water and sewage treatment processes in removing or destroying viruses and bacteria in water or sewage. Kabler et al.[5] summarized the viricidal efficiency of a number of water disinfectants. They concluded that different types of enteric viruses vary widely in their resistance to chlorine and that at least some types of polio-virus, Coxsackie viruses, or E.C.H.O. viruses are more resistant to chlorine than are coliform or enteric pathogenic bacteria. Although some of these viruses are markedly more resistant to chlorine than are coliform bacteria, Clarke and Chang[6] concluded that any enteric virus so far examined would be effectively destroyed by a free chlorine residual of about 1.0 ppm, provided it could be maintained for about 30 minutes and the virus particles were not embedded in particulate material....

"There is no doubt...that water can be adequately treated so that it is always biologically safe. Adequate treatment means good clarification...followed by effective disinfection which can be carried out only with water free from suspended material. This has been pointed up very clearly by Sanderson and Kelly[7], who cite a case in which coliform bacteria were constantly isolated from waters containing between 0.1 and 0.5 mg/l of free residual chlorine and between 0.7 and 1.0 mg/l of total residual chlorine after 30 minutes contact time. They also point out that data on the effectiveness of chlorine disinfection as determined by laboratory studies conducted under favorable conditions are not applicable to turbid waters. The chlorinated water in which they consistently found coliform organisms had turbidities ranging from 3.8 to 8.4 units, had iron rust, and had biological organisms of 2000 units on occasion. Viruses, being much smaller than bacteria, would appear to have more opportunity to become enmeshed in a protective coating of turbidity-contributing materials and thus more opportunity to escape disinfecting action. It is therefore advisable that turbidities in final effluents be kept well below one Jackson unit. In fact, for most effective chlorination it would be well to have the turbidity as low as 0.1 unit as recommended by A.W.W.A. water quality goals. (Incidentally, it should be understood that the limit of 5 Jackson Units of turbidity specified in the Public Health Service Drinking Water Standards of 1962 is meant to apply to protected watersheds and not to effluents from filtration plants.) With turbidities as low as 0.1 to 1, a pre-plant chlorine feed need be only enough to have a 1 mg/l free chlorine residual after a contact time of 30 minutes. Post chlorination practice would depend upon the ability to maintain such residuals throughout the distribution system."

The A.S.C.E. Committee then goes into the following conclusion:

"There is no doubt that a positive coliform index means that virus may be present. Absence of coliform, however, may not mean that virus is absent. The coliform index, therefore, while a good laboratory tool is not a reliable index for viruses. Greater assurance of the absence of virus would be a turbidity of less than 0.1 Jackson Unit and an HOCl residual of one ppm after a contact period of 30 minutes. While freedom of virus hazard may be associated with the latter conditions, it would be desirable that a viral index be utilized to supplement the coliform index. Virologists engaged in water supply research should consider this need."

We have a good laboratory tool for the determination of chlorine residuals in the amperometric titrimeters. We lack a good tool for the measurement of turbidities in the range of 0.01 to 1.0 Jackson Units. If such a reliable instrument were made available it would be possible to commence studies on the control of infectious bacteria and viruses by relating them to the chlorine residuals and turbidities.

Conclusions

The very first Drinking Water Standard Committee added the following concluding comment to their very important work. "It is requested that the recommendation of these hard-and-fast limits of bacteriological impurity be not interpreted as minimizing in any way the importance of field surveys in estimating the sanitary quality of water supplies in general. It is always desirable to obtain information from as many angles as possible, and this is, indeed necessary in order to form an altogether fair estimate of an individual supply." This is precisely the same situation we are in with respect to the domestic reuse of wastewaters, viz., "It is...desirable to obtain information from as many angles as possible..."

While studies of bacteria and virus pathogen control should be directed to HOCl residuals, time, and turbidity, we will for a long time have to couple these analyses to viral assays, coliform count, and, indeed, include other possible microbiological problems, perhaps using a total count procedure and/or specific quantitating techniques for such organisms as pseudomonads which have been reported in many instances to be problems with several tertiary treatment devices.

In 1919, a Public Health Service medical officer stated[8]: "Water supplies heretofore have been passed upon by bacteriological standards. Very little attention has been paid to (sanitary) survey standards which engineers would likely set. I think the time has come to adopt the engineering point of view. The bacteriologist will only be a checker-up and I think I can convince you... that we should adopt the (sanitary) survey method of accepting a water supply rather than the bacteriological."

I submit that this is a big world with plenty of room for engineers, bacteriologists, virologists, chemists, and medical doctors. Certainly it will take an abundance of teamwork involving all disciplines to achieve bacteria and viral control in water reuse. The Public Health Service and the Committee on Environmental Quality Management of the Sanitary Engineering Division, A.S.C.E. have clearly pointed the way.

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